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National Environmental Research Center
Cincinnati, Ohio

May 1975

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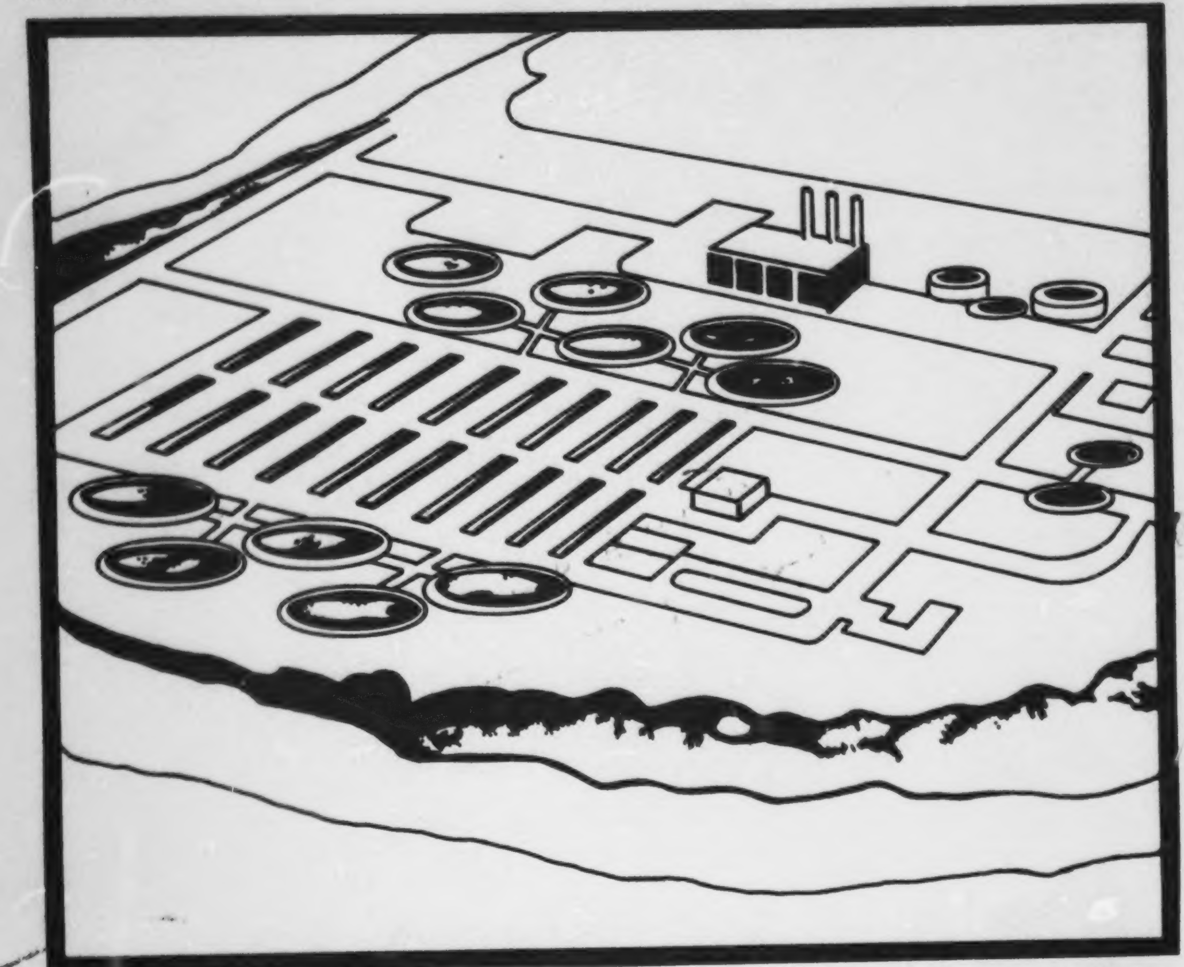

PROCEEDINGS

THIRD U.S.-JAPAN CONFERENCE ON SEWAGE TREATMENT TECHNOLOGY

February 12-16, 1974
TOKYO, JAPAN

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At the Third U.S.-Japan Conference on Sewage Treatment Technology, the Japanese delegation presented a series of papers expressing views on the Federal Water Pollution Control Act Amendments of 1972, specifically of the water quality and effluent standards and grants for construction of treatment works; detailing research done and problems encountered in the heat treatment of sewage sludge; outlining the feasibility of combined treatment of municipal and industrial wastewater; and reporting the status of studies being done on advanced waste treatment. Members of the U.S. delegation presented State and Federal viewpoints on the Federal Water Pollution Control Act Amendments of 1972; the overall EPA wastewater treatment research program; sludge treatment and disposal guidelines for municipal wastewater plants in the United States; methods of wastewater treatment by physical-chemical nitrogen removal; experiences with sludge handling and oxygen activated sludge systems in Texas; studies on aeration systems, suspended solids removal processes, and nitrification conducted in Metro Chicago; and EPA experiences in oxygen activated sludge.

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THIRD
U.S. - JAPAN
CONFERENCE ON
SEWAGE TREATMENT TECHNOLOGY
PROCEEDINGS

February 12-16, 1974
Tokyo, Japan

OFFICE OF INTERNATIONAL ACTIVITIES
OFFICE OF RESEARCH AND DEVELOPMENT
OFFICE OF WATER AND HAZARDOUS MATERIALS

U.S. ENVIRONMENTAL PROTECTION AGENCY
CINCINNATI, OHIO

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FOREWORD

The United States and Japan share a common concern for the protection of man's environment. Both nations have recognized that their highly developed technological talents should be turned to the solution of environmental problems which confront us today. In the past they have made important advances, both individually and in cooperation with others, to preserve and enhance the quality of life.

The advantages which accrue from cooperative effort on problems of mutual concern are undeniable. Building on two earlier successful meetings, the Third U.S./Japan Conference on Sewage Treatment Technology, held in Japan in 1974, is a recent example of effective cooperation for mutual benefit. These Proceedings will be of real value to future efforts in the field.

We look forward to similar productive cooperation in the future.

Russell E. Train
Administrator

Washington, D.C.
April, 1975

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JOINT COMMUNIQUE

THIRD U.S./JAPAN CONFERENCE
ON
SEWAGE TREATMENT TECHNOLOGY

Tokyo, Japan
February 16, 1974

1. The Third U.S./Japan Conference on Sewage Treatment Technology was held in Tokyo from February 12-16, 1974 by mutual agreement between Mr. Russell E. Train, Administrator of the U.S. Environmental Protection Agency (EPA) and Mr. Takao Kameoka, Ministry of Construction.

2. The U.S. Delegation headed by Mr. Francis M. Middleton, Deputy Director of The National Environmental Research Center (NERC), Cincinnati, Ohio, EPA was accompanied by seven representatives from the U.S. EPA, state and local governments.

3. The Japanese delegation headed by Dr. Takeshi Kubo, Director General, Department of Sewerage and Sewage Purification, Ministry of Construction, was composed of eight National officials, three local government officials, and two representatives from the Japan Sewage Works Corporation.

4. Prior to the Conference the U.S. delegates visited Saitama Prefecture Arakawa Sewage Treatment Plant, Tokyo Metropolitan Government Ukima Works and Ochiai Works, Yokohama City Torihama Industrial Waste Treatment Works, Yokosuka City Shitamachi Sewage Treatment Plant,

Zushi City Water Pollution Control Center, Fujisawa City South Sewage Treatment Plant, Atami City Nishikigaura Sewage Treatment Plant, Kyoto City Toba Sewage Treatment Plant and Kisshoin Sewage Treatment Plant, Nara Prefecture Yamato River Purification Center and advanced waste treatment pilot plants at Yokosuka and Kyoto which were conducted by the Ministry of Construction.

5. Each field visit involved the subject matter to be discussed during the Conference.

6. The U.S./Japan Conference on Sewage Treatment Technology grew out of the Second U.S./Japan Ministerial Conference on Environmental Pollution held at Washington, D.C. in June, 1971 between Chairman Russell E. Train, then Head of the Council on Environmental Quality and then Japanese Minister Sadanori Yamanaka. The First Conference was held at Tokyo, Japan in 1971 and the Second Conference was held at Washington, D.C. in 1972.

7. Principal topics of the Third Conference were Federal Water Pollution Control Act Amendments of 1972, sludge treatment and disposal, combined municipal and industrial waste treatment and advanced waste treatment technology. During the Conference the presentations were followed by vigorous discussions from both sides.

8. In addition to the Conference, the Municipal Design Seminars were opened to about two hundred members of the Japan Sewage Works Association on the subjects of physical-chemical treatment technology, pure oxygen activated sludge technology, upgrading of existing plants, and combined sewer and stormwater technology.

9. A video tape concerning research on environmental pollution at the Robert A. Taft Center, NERC, EPA, Cincinnati, Ohio was narrated in Japanese

by Mr. Tokuji Annaka, research engineer, Public Works Research Institute, Ministry of Construction. Mr. Annaka is spending one year in Cincinnati by an exchange program of technical personnel between two countries in a follow-up of conclusions reached at the Second Conference held at Washington, D.C. in 1972.

10. The Conference was successful and fruitful in the exchange of knowledge and experience with each country. Future emphasis will be placed upon closer exchange of experts and information between the two countries.

A progress report of these three Conferences since 1971 will be presented by both sides for the coming U.S./Japan Ministerial Conference on Environmental Pollution.

11. The delegations agreed to explore and identify research projects, which might be undertaken jointly by both U.S. and Japanese experts, and also to look for the possibility of attendance of officials from Asian countries to the future conferences.

AGENDA

JAPANESE PRESENTATIONS

TUESDAY, FEBRUARY 12:

JAPANESE SIDE VIEWS ON U.S. FEDERAL WATER POLLUTION CONTROL ACT AMENDMENTS OF 1972

WEDNESDAY, FEBRUARY 13:

HEAT TREATMENT OF SEWAGE SLUDGE

COMBINED TREATMENT OF MUNICIPAL AND INDUSTRIAL WASTEWATER

THURSDAY, FEBRUARY 14:

STUDIES ON ADVANCED WASTE TREATMENT

JAPANESE SIDE VIEWS ON THE FEDERAL WATER POLLUTION CONTROL ACT AMENDMENTS OF 1972

WATER QUALITY STANDARDS, EFFLUENT STANDARDS

presented by
Katsumi Yamamura
Head, Water Quality Control Division,
Water Quality Bureau,
Environment Agency

GRANTS FOR CONSTRUCTION OF TREATMENT WORKS, RESEARCH AND RELATED PROGRAMS

presented by
Takeshi Kubo
Director, Department of Sewerage & Sewage Purification,
City Bureau,
Ministry of Construction

February 12-16, 1974

Ministry of Construction
Japanese Government

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1. Water Quality Standards, Effluent Standards

As the contents and intentions of the Federal Water Pollution Control Act of 1972 were made clear through the Guidelines and other informations recently issued, we, as Government officials responsible for water pollution control, highly appreciate such informations and the Act itself as good references for considering the measures to combat water pollution such as establishment of regulatory standards and as basic informations for reviewing the measures already taken.

First, we feel that the FWPCA of 1972, which most comprehensively provides for the policy to protect and improve water quality, is more functional than the statutory system in Japan, composed of several laws i.e. the Basic Law for Environmental Pollution Control which is the basic law for pollution control policy in general and provides for the establishment of Water Quality Standards, and the Water Pllution Control Law which provides for various regulatory measures and other laws to carry out pollution control projects including construction of sewerage.

The second point I want to say is on Water Quality Standards. The Water Quality Standards under the FWPCA of 1972 are to be established to attain a very high quality of the environment so as to protect fishes and other aquatic living organisms and their culture, and to maintain the natural beauty and the amenity for peaple's recreation activities. We would highly appreciate such intensive water pollution control policy when considering its difficulty in attaining such high quality in any water bodies which are already polluted to some extent including closed waters such as harbors around industrial zones.

The third point is on effluent standards. I think that you have made great change in ways of thinking and those provisions on best practicable or best available technology are to be applied in all bodies of water. This would give great influence on water management policy in Japan.

According to the informations already published, the effluent standards have been set as maximum emission units e.g. maximum COD emission amount per unit raw material in cases when best practicable or best available technologies are adopted. We have not been able to examine the effluent standards in detail but we have found that some of them are very stringent. We think that, in order to establish such standards, you had to consider current effluent treatment technologies, differences in manufacturing processes, possible effects on production costs, etc., and we highly appreciate the efforts you have made.

I would like to ask you a few questions on the FWPCA of 1972, the Guidelines and related matters.

(1) Any State Governor has to consult with the Administrator of EPA when he wants to work out a plan for pollution control, development and use of water resouces (including restoration, conservation and protectio of them) with the view to preventing water pollution. I want to know the methods and procedures of the consultation and environmental impact assessment between State Governor and the Administrator.

(2) You have established to effluent standards as emission amount of pollutant per unit raw materials. I think that continuous manitoring systems of quality and quantity of effluent and of consumption of raw materials should be established. How do you carry out such monitoring?

(3) What like are the cases when the establishment of least water quality standards is considered impossible because of natural conditions of water, man-made pollution and special purpose of use of water? I think that the acidic waters caused by mines is one of those cases. How do you establish water quality standards in this case? And do you carry out any measures to improve the quality of such acidic waters?

(4) Some bottom deposits may give bad influence on fishes, ecosystem and water quality. Organic deposits will influence on eutrophication.

For example, mercury in fishes of Minamata Bay is considered to be caused by contaminated deposits. In order to combat pollution by mercury and its compounds, we established the Provisional Standards for Removing Bottom Deposits Containing mercury on June 30, 1973. Do you have any intention to establish standards for removing bottom deposits and to enforce them?

2. Grants for Construction of Treatment Works, Research and Related Programs

2.1 Introduction

Environmental pollution control has been one of the important policies recently in Japan especially for these several years and it is further recognized that the policy of the environmental pollution control shall be placed on a top priority among various national policies from now on. Because, initially the pollution problem was attacked on a fragmentary basis with various local governmental units coped with the independent countermeasure for special types of pollutant, but the environmental problems, particularly water pollution problems relating to rivers, lakes and coastal waters are accused by the public not only in the industrial area of the large cities, but also in the rural area of all over Japan, and so the public has become more aware of water pollution and its attendant environmentally degrading problems in such a way of the fishermen's union's demonstration against pollution in various part of Japan. Under these circumstances the positive protective actions have been seriously demanded with increasingly periodicity.

Japanese Government has been trying to establish powerful legislation on pollution control and also to promote the countermeasure against water pollution prevention such as sewage treatment construction planning. This is my understanding that the Federal Water Pollution Control Act Amendments of 1972 (the Act) is a very comprehensive and complex law that addresses all types of water pollution, and so far in Japan we could not go through such a experience on its comprehensiveness and its ambitious national goals. In the beginning of the Act the national goals are declared in taking up a positive attitude. The ultimate goal is to eliminate the discharge of pollutants into the navigable waters of the United States by 1985 and the interium goal is to achieve the water quality by July 1, 1983 which provides for the protection and propagation of fish, shelfish and wildlife and provides for recreation in and on the United States waters. Further, the national policy is that the discharge of toxic pollutants in toxic amounts be prohibited. The concrete measure and the definite procedure to attain the goal can be realized through the regulations and guidelines pursuant to each section of the Act.

It is understandable that the regulations establish the terms of requirements for each specific piece of the Act and the guidelines explain, amplify in detail, and provide guidance to supplement the regulations, and in these connections the regulations are formal documents that provide the legal framework for the implementation of the Act and on the other hand the guidelines are explanatory and assist in the understanding of the regulations. I have gotten through with reading some regulations and guidelines and I would like to present the Japanese side views and to make some comments on the Act, regulations and guidelines.

2.2 Target Dates

In order to achieve the ultimate national goal you have only ten years more. With full implementation of the Act including target dates for the use of "best practicable" technology subsquently "best available" technology, and finally zero discharge of pollutants, and in these connections you shall have to carry out the extraordinarily large waste treatment works program. It seems to me that it takes time to explore to make the most reasonable planning and also designing with waste treatment facilities from view point of engineering, and as might be expected in such a development, by now it is highly doubtful that any local government would meet its responsibility to control water pollution unless the maximum authorized federal grant is provided, and the 18 billion dollars federal grant program will be increasing due to construction cost rising, and also there must be many difficulties to cope with the time limitation of the Act for small communities and small industries. From these points of view I am obliged to acknowledge that it will be too short to attain the target dates for the ambitious national goals.

2.3 State Grants for Construction of Treatment Works

Before enactment of 1972 Amendments Act, the program was aimed at providing an incentive for State Governments and local municipalities and offered 10% bonus federal grant in the case of combination of state grant program and of regional waste treatment plant construction. It seems that such an incentive policy would be quite effective to promote construction program with each financially sharing basis among Federal, State and local municipality. The national goals of the Act should be the mutual goals of all concenrned with protection of U.S. Nation's waters and in order to achieve the national goals it should be absolutely necessary to keep a coordinate effort with each other among EPA, State agencies, municipalities, design engineers, and industries. In order to keep such a partnership some kinds of incentive with each other would be desirable. The current program promises seventy-five percentage federal grant of the construction cost of sewage treatment plants and sewers for publicly owned facilities of any size and in this case there is no incentive between EPA and state agencies. I would say that collaboration and cooperation should be maintained at the federal, state and regional level. Through the experience in the past Act is there any bad effects in such an incentive way of federal grant system between federal and states.

2.4 Reactions for Guidelines from Professional Engineers

It seems that the regulations and guidelines establish a very powerful frame work for the purpose of establishing minimum requirements to be followed by all planners and designers of municipal water pollution control facilities, and I can imagine that there must be the tendency of guidelines to be used as standards. The objective of the Act is to restore and maintain the chemical, physical and biological integrity of the Nation's waters and this tendency should not cause the regulations and the guidelines to become too confining, because this tendency could stifle the creativity of the professional designers. I think that standardization on design criteria of facilities sometimes will give a bad effect to creative improvement from view ponts of engineering. Do you have any reactions or arguments from the professional designer's field on the matters of standardization caused by establishing the regulations and guidelines of the Act?

2.5 Secondary Treatment Information

Pursuant to section 304 (d) (1) the information, in terms of the parameters of biochemical oxygen demand, suspended solids, fecal coliform bacteria and pH on the degree of effluent reduction attainable through the application of secondary treatment. Terms '1-week' and '1-month' as used in effleunt samples are expressed in a period of seven consecutive days and thirty consecutive days. It is reasonable in this way to have the parameters of BOD, SS and fecal coliform bacteria. It is understandable that in the case of combined sewer system during wet weather requirements should be made case by case, but it seems to me that there should be some guidelines to make the decision of the attainable percentage removal level in combined sewer during wet weather. In most of large cities in Japan the combined system has been taken and many storm-overflows do exist exactly on sewers. In these cases it is not so practicable to decide the requirements even in case by case. I understand in the regulations that when industrial wastes discharge into publicly owned treatment works, effluent standard can be adjusted upwards to only those cases in which the flow or loading from an industrial category exceeds 10 percent of the design flow or loading of the treatment works. Actually the percentage of the flow or loading of industrial waste into publicly owned treatment works may vary from time to time, and this means that the effluent standards from the corresponding publicly owned treatment works may vary from time to time being adjusted proportionally. Who is taking responsibility to decide the effluent standards for such a case by case such as combind sewer and industrial waste problem? I shall be much obliged if someone explain the practice in U.S.A. more in detail for combind sewer and industrial waste.

2.6 Information on Alterrative Waste Treatment Management

Pursuant to section 304 (d) (2) of the Act, information on alternative waste treatment management techniques and systems available to implement section 201 of the Act, is published in dividing into three categories (1) treatment and discharge into navigable waters (2) land application techniques and (3) wastewater reuse. In navigable waters, where water quality standard are more stringent, effluents based on those standards will apply and more stringent effluent limitations including additional parameters such as ultimate oxygen demand and ultimate biochemical oxygen demand will be decided. The use of the parameters of ultimate exygen demand means the total oxygen demand of the wastewater effluent including organic nitrogen, especially in the form of ammonia. I seems to me that technology to satisfy the ammonia biological demand is more costly and less widely achievable and is not always practicable at the moment. I wonder whether this kind of technology should be called to be best practicable or not.

2.7 Infiltration/Inflow

Pursuant to section 201 (g) (3) of the Act, the EPA Administrator shall not approve any grant after July 1, 1973, for treatment works under this section unless the applicant shows to the satisfaction of the Administrator that each sewer collection system discharging into such treatment works is not subject to excessive infiltration. It is recognized that sewage treatment of infiltration/inflow requires larger treatment works with increased cost for capital, operation and maintenance, and elimination of infiltration/inflow by sewer system rehabilitation can often reduce the cost of sewage collection can often reduce the cost of sewage collection and treatment, therefore a logical and systematic on the total system over sewage collection and treatment is necessary to determine the cost-effectiveness through the cost-effective analysis. The cost-effective analysis should be based on the comparison of the estimated cost for treatment of infiltration/inflow and the estimate cost for rehabilitation of the sewer system. It seems to me that in this case the estimate cost for treatment of infiltration/inflow may vary sharply in either case of estimating the cost for best practicable or best available process, in other words for secondary treatment or advanced waste treatment.

2.8 Allotment of Federal Grants for Construction of Treatment Works

As for allotment of federal grant and state determination and certification of project priority, pursuant to section 205 (a) the federal grant of plant construction shall be allotted among the States in accordance with regulations in the ràtio that the estimated cost of constructing all needed publicly owned treatment works in each state bears to the estimated cost of construction of all needed publicly owned treatment works in all of the states. I understand that pursuant to section 516 (b) (2) a detailed estimate, biennially revised, of the cost of construction of all needed publicly owned treatment works in all of the States and of the cost of construction of all needed publicly owned treatment works in each of the States and the applicable percentages to be used in computing State allotment every two years. On other hand the EPA Administrator, pursuant to section 516 (b) (3) (4), shall make a comprehensive study of the economic impact on affected units of government of the cost of installation of treatment facilities, and a comprehensive analysis of the national requirement for effluent to attain water quality objectives as established by the Act. Is there any possibility for the Administrator to correct the allotment of funds at the results of the comprehensive study and analysis pursuant to section 516 (b) (3) (4)? Pursuant to the regulation of grants for construction of treatment works, construction grants will be awarded from allotments available in accordance with a State System for certification of priority for construction grant project. What kind of criteria is used in Texas State priority system?

2.9 Public Participation

It is required by the regulation that the State agency has afford adequate opportunity for the public participation such as oral or written municipal and public comment upon the State priority system. Further, pursuant to section 101 (e), the EPA Administrator, in cooperation with the States, shall develop and publish regulations specifying minimum guidelines for public participation in such processes. Could you explain the public participation in the development of water pollution control program in the case of Texas State agency?

2.10 Cost-Effective Analysis

In cost-effective analysis procedure of the guidelines pursuant to section 212 (2) (c) of the Act it is required that the planning period for the cost-effective analysis shall be 20 years. On the other hand pursuant to section 208 (b) (2), any plan prepared under such process shall include the idenfication of treatment works necessary to meet the anticipated municipal and industrial waste treatment needs of the area over a twenty years period including any requirements for the acquisition of land for treatment purposes; the necessary wastewater collection and urban storm runoff systems; and a program to provide the necessary financial arrangements for the development of such treatment works. I wonder why a twenty-year period is fixed in those cases, because pursuant to the same guidelines of cost-effective analysis the service life shall be taken in accordance with such situations as follows;

Land	permanent
Structures	30 ~ 50 years
Process equipment	15 ~ 30 years
Auxiliary equipment	10 ~ 15 years

2.11 Pretreatment Standards

It seems that the federal pretreatment standards pursuant to section 307 (b) of the Act are intended to be national level in scope and in many cases it will be necessary for a State or a municipality to supplement the Federal Standards with additional pretreatment requirements in accordance with the local condition pursuant to section 304 (f) (1) and guidelines. Pretreatment for removal of compatible pollutants such as BOD, SS and so on, is not required by the Federal Pretreatment Standards. I understand this indicates that the compatible industrial wastewater will be encouraged to be treated in publicly owned treatment works by joint treatment under the user charge systems. I understand also that under the National Pollutant Discharge Elimination System pursuant to section 402, all point sources including publicly owned treatment works must obtain a permit for the discharge of wastewaters to the navigable in the United States, permits, however, will not be required for industrial sources discharging into publicly owned treatment works, and effluent limitations for publicly owned treatment works is required including (a) secondary treatment information, section 304 (d) (1) (b) toxic effluent standards, section 307 (a) (c) Water quality standard, section 303 of the Act, and the most stringent limitation for each pollutant will govern. Accordingly I understand that it may be necessary to establish a local system which will allocate waste loads to industrial users so that biological treatment processes are not inhibited and to ensure that effluent limitations are met. Finally there is one thing in pretreatment for the incompatible pollutants such as heavy metals. In the case of pretreatment to remove heavy metals there must be some difference on quality limitation between publicly owned biological treatment works and physical chemical treatment works. Supposing that small amount of heavy metals through legal pretreatment limitation are discharged into publicly owned treatment works and such heavy metals will be concentrated in the sludge, and there will happen to cause difficulties in sludge treatment and disposal. In this connection there are serious discussions in Japan in which any heavy metals in any amount into publicly owned treatment works should be prohibited from industrial sources including small industries, and publicly owned treatment works should not take responsibility of recieving any heavy metals from industry.

2.12 User Charge

It seems that the user charge systems are intended to enable the treatment authority to be financially self-sufficient with respect to operation and maintenance, because the term of operation and maintenance in this case includes replacement, and the expenditures for replacement are the expenditures for installing equipment or appurtenances which are necessary during the service life of the treatment works to maintain the capacity and performance for which such works were designed and constructed, and also user charges must be included in the charges made by treatment authorities for wastewater collection and treatment such as payments for local debt service for previous construction and local share of the works. I understand that the user charge system may be a policy under which sewage treatment works should be financially self-suport excluding reconstruction and expansion works.

As for industrial cost recovery, it seems that the definition of an industrial user is too broad, especially with regard to Division I - Service - of the Standard Industrial Classification Manual, 1972, Office of Management, because some kinds of service industries which are having close contact with civic life discharge primarily domestic type waste, and are different from industries contributing significant quantities of process wastes. In Japan discussions are now going on with regard to the definition of the industrial user particularly relating to service industry, because we are going to establish the specifyed sewer charging system for industrial users in accordance with the pollutors pay principle, ppp. Finally I would like to raise questions concerning with treatment of stormwater and infiltration. How the cost for stormwater treatment and infiltration water treatment will be financed? In our Japanese practices user charges are levied by the way of system based on water quantity measured by water meters installed in each house, and of course such water quantity does not include stormwater and infiltration water, and we cannot distribute such portion of such cost of stormwater and infiltration water treatment to users.

2.13 Industrial Cost Recovery

The regulations require that all grantees recover from industrial users that portion of the grant amount of the treatment of wastewater from such users, and an industrial users' share shall not include an interest component. The regulation also provides that a grantee may retain an amount of the revenues recovered from industry equal to (1) the amount of the non-Federal cost of the project paid by the grantee, plus (2) the amount necessary for future reconstruction and expansion of the project. The total amount retained, however, cannot exceed 50 percent of the amount recovered. There are three points arising here that the first one is the reason why 50 percent limitation in retainment is taken and the second one is the fact that the corresponding treating authority may make a profit in this portion through industrial recovery system, and the third one is that there must be arguments in which industrial users may not pay the interest component, but other industries which do not happen to discharge their wastewater into public sewers due to their location cannot help providing their own plants and operating the plants by their own expenditures including the interest component.

2.14 Consideration on Reclaiming or Recycling of Water

Pursuant to section 201 (g) (2) (B) provides that the works proposed for grant assistance will take into account and allow to the extent practicable the application of technology at a later date which will provide for reclaiming or recycling of water. At the planning or designing basis what kind of consideration should be definitely paid for reclaiming or recycling of water?

2.15 Planning of Storage for Water

Pursuant to section 102 (b) (1) (2) (3) (4) (5) of the Act, in the planning of storage for water quantity and regulation of stream flow consideration shall be given by the Corps of Engineers, Bureau of Reclamation, or other Federal agency. The Act provides also that the need for value of storage for regulation of stream flow shall be determined by the Corps of Engineers, Bureau of Reclamation and the value of storage for quality control shall be determined by the EPA Administrator. It seems that the program of quality control and the program of quantity control should not be separated and should be unified at least in the planning basis. It is reported that in Britain reorganization in the field of water service is going to be established and 10 New Regional Water Authorities will be established in England and Wales, and they will take responsibility of all water problems including quantity control, regulation of stream flow, water pollution control, water supply, sewage purification and disposal and so on. The regional water authority which will be established in its river basin wide base is really single organization for water quality and quantity in all water problems.

The British practice in this way may be of some value for our discussion.

HEAT TREATMENT OF SEWAGE SLUDGE

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HEAT TREATMENT OF SEWAGE SLUDGE

1. Introduction

1.1 Antecedents

As of 1972, 254 sewage plants were operated in Japan, and the number now is on the constant increase. Those plants were running mostly on a method embracing a series of processes including anaerobic digestion, mechanical dewatering, dumping or incineration. As for the dewatering process, the mechanical method seems to have become more increasingly practised in recent years. This is primarily because the process resting on anaerobic digestion, drying with bed and dumping is being discouraged by an offensive taken by nearby inhabitants against nuisance stink, difficulties in acquiring suitable plant sites, labour shortage, etc.

On the other hand, as more and more the mechanical dewatering process has been disseminated, its characteristic problems have come to light.

The following is a list of major problems and a breif explanation of each.

1) Economics

The mechanical dewatering process requires a pretreatment in which coagulant is dosed to thickned sludge for increasing dewatering efficiency.

In Japan, combined use of ferric chloride ($FeCl_3$) and lime is the most commonest of all as a coagulant dose. In some cases, however, ferric sulfate ($FeSO_4$) is used instead of ferric chloride. Anyway, the chemicals accounts for as much as 20 to 45% of the total operating cost of the sludge treatment.

2) Working efficiency

It is often that the content of organic substances reaches no less than 80% or more in the sludge to be handled at sewage plants of large housing communities, local cities and especially those designed as a separate sewer system conveying mostly domestic sewage. This kind of sludge is very fine in size, and is hard to cencentrate; if it is retained in a thickener for 2 to 3 hours, it will readily come afloat as scum, and worse the moisture content of the cake cannot be kept at the level of 75 to 80% unless the plant operator doses more coagulants than usually required.

Increase in the content of lime in the sludge cake increases sludge volume as a whole, over loading the incinerator.

Since the separate sewer system is expected to gain popularity in the future, the above problems will emerge as a reality.

3) Working conditions

The vacuum filters, filter presses and other machines widely used for the mechanical dewatering process cannot do without constant inspection, cleaning and overhaul, and the operators are forced to attend them all the time. Meantime, the working environment in the sludge treatment process is insalubrious compared to the rest of the sewage treatment system. Lime handling is really a peeve, because this fine powder is liable to fly about.

As a consequence, it poses grave concern over labour settlement and occupational hygiene to add to the fact that the sludge treatment is an extraordinarily costly business what with upkeep, maintenance, inspection and soaring labour cost.

4) Durability of equipment

Ferric chloride used as a coagulant is very active upon metal structures such so chemical storage tank, mixing tank, piping, valves, gas ducts and fans in incinerator, costing replacement and repair too much. The improvement in the materials of such equipment is a matter of primary concern.

To cope with these and other various problems, emphasis has been placed on the development of new systems with which to increase dewaterability and realize remote and automated control.

To ward off the problems resulting from the use of chemicals, heat treatment process has attracted keen attention as a promising pretreatment method of sludge. In Japan, this process was considered for the first time in 1970 when the modification of a sludge handling system of Shyojaku Plant was projected, and eventually was practised after investigation of a test plant.

In this process, sludge is heated in order to coagulate protein in it into hydrophobic one, remove bonding water as well as to improve sedimentability and dewaterability by accelerating the flocculation of suspended solids. It is evident that the process will save installation space and cut labour by automation. The process is also favoured from the viewpoint of sanitation that the heating annihilates baccili and parasite eggs.

The Ministry of Construction, while believing in the process, urged to appraise it from various angles of view, turn up problems and provide measures against them before implementation as it was concerned about the safety and durability of the equipment and offensive byproducts such as stink and high concentrate supernatant.

To push forward the Ministry's policy, full-scale plants were installed in Sakai (Semboku Plant), Fujisawa (Nambu Plant) and Sapporo (Toyohiragawa Plant), so that the performance of the plant facilities could be assessed. Also, the Committee for Investigation into Sludge Handling and Disposal of the Japan Society of Civil Engineers was entrusted with the task of conducting fundamental studies, evaluation of plant achievements and retrieval of problems. Realizing the significance of the mission assigned to it, the Committee immediately organized a subcommittee comprising civil engineers, sanitary engineers, metallurgists, mechanical engineers, and of course sewage engineers from the municipalities operating the full-scale plants, and embarked on research work from 1970.

Their principal subjects were:

1) Literature research on the process achievements in European countries

2) Laboratory test on sludge available here in Japan to elucidate the principles, effects and problems of the heat treatment process

3) Assessment of full-scale plants based on actual operation data from the viewpoint of economics, durability of facilities and operatability, etc.

4) Examination of characteristics of supernatant and development of its treatment techniques

5) Detection of stink factors and examination of methods for removing them

6) Research of the equipment with respect to corrosion and banking, etc.

These surveys were actively promoted in both field and laboratory. The proceedings were made public annually in the form of interim report, and some of them were submitted to the Second U.S.-Japan Conference on Sewage Treatment Technology.

Submitted herewith is the summary of the final report covering the three-year research activities of the committee and its conclusions.

1.2 Laboratory tests on heat treatment

In corroboration of the principles of heat treatment and for the purpose of obtaining the optimum treating conditions, fundamental laboratory tests were conducted using sludges from Semboku Plant, Nambu Plant and Toba Plant, Kyoto.

The results were as follows.

1) Both excess activated sludge and sludge from the primary sedimentation tank tended to be filtered quickly when treated at temperatures above 180°C. The higher the temperature, the lower the moisture content in the cake. The heat treating time was not a significant factor for both sludges, but with the temperature and time fixed, the sludge from the primary sedimentation tank could create the cake of lower moisture content than the excess activated sludge did.

2) The higher the heat treating temperature and the longer the heat treating time, the faster the filtration rate. With the treating time fixed, the filtration rate increased in proportion to the temperature.

3) While the specific resistance of sludge had no significant correlations with heating temperature and time, if up to 180°C, as it was largely affected by the content of organic substances, it plunged down at temperatures above 190°C.

4) For both excess activated sludge and sludge from the primary sedimentation tank, BOD, COD and ammonia nitrogen in supernatant increased sharply from temperatures 180°C and above; their concentrations became higher on the higher temperatures and on the longer time. Similar tendency was noticed with respect to the colour of the supernatant.

5) The solubility of heat treated sludge was higher as the organic content increased; namely, the highest was the excess activated sludge, followed by mixed sludge and sludge from primary sedimentation tank in turn. It was inferred that some 50 per cent of solids contained in the activated sludge would dissolve in some 30 min. of heat treating.

6) The baking of heat exchanger become violent when the heating temperature was increased more than 200°C. The tendency for the heat exchanger to have scorches was stronger as the organic content increased.

7) It was concluded that the heat treatment as a pretreatment of sludge for dewatering can be carried out effectively even with temperature set at 180°C and time at 30 to 60 min. if the organic content in sludge is in the range of 50 to 60%, while if the organic content varies or exceeds 60%, the heat treating conditions should be set at 190°C to 200°C and 30 to 60 min.

Sewage treatment plants where full-scale facilities were installed are listed in Table 1.1. They all were designed to handle for the most part domestic sewage.

Table 1.1 shows the design data for each of the plants. Their heat treatment facilities are outlined in Table 1.2, and the process flow of sludge treatment section is given in Fig. 1.1.

Table 1.1 Design data for heat treatment facilities in projected sites

Items \ Plant	Toyohiragawa, Sapporo	Nambu, Fujisawa	Semboku, Sakai
Date of operation	Oct., 1970	Aug., 1964	Mar., 1969
(1) Sewer system	Combined sewer (domestic sewage only)	Combined sewer (partly combined & partly incl. industrial effluents)	Separate sewer (domestic sewage only)
(2) Description			
Plant area	8.557 ha	9.24 ha	8.40 ha
Served area	2,202 ha	1,713 ha	1,845 ha
Served population	200,000	228,000	221,000
Volume of sewage flow, avg.	64,000 m^3/day	76,730 m^3/day	86,190 m^3/day
(3) Sewage treatment process	Activated sludge process (step aeration possible)	Activated sludge process (step aeration system)	Activated sludge process (step aeration system)
(4) Heat treatment system	Porteous system	Von-Roll system	Porteous system

Plant Items	Toyohiragawa, Sapporo	Numbu, Fujisawa	Semboku, Sakai
(5) Design data			
1) Quality of influent	B.O.D. 200 ppm S.S. 250 ppm	B.O.D. 200 ppm S.S. 250 ppm	B.O.D. 200 ppm S.S. 300 ppm
2) Total solid of raw sludge	20.16 t/day	5.8 t/day	7.84 t/day
3) Moisture content of raw sludge	99%	98%	98%
4) Volume of raw sludge	2,345.3 m^3/day	288 m^3/day	392 m^3/day
5) Moisture content of thickened sludge	95%	96%	96%
6) Volume of thickened sludge	403 m^3/day	144 m^3/day	196 m^3/day
7) Moisture content of heated sludge	90%	92%	90%
8) Total solid of heated sludge	16.12 t/day	4.6 t/day	5.84 t/day
9) Volume of heated sludge	161.2 t/day	57.7 t/day	54.9 t/day

Plant Items	Toyohiragawa, Sapporo	Numbu, Fujisawa	Semboku, Sakai
10) Moisture content of cake	47%	40%	50%
11) Cake volume	30.4 t/day	7.7 t/day	9.97 t/day
(6) Heat treatment facilities and operating conditions			
1) Raw sludge temperature	10°C	20°C	15°C
2) Outlet temperature of heat exchanger	165°C	160°C	160°C (from reactor)
3) Heat treating temperature in reactor	200°C	200°C	200°C
4) Outlet temperature of heat exchanger	55.3°C	62°C	60°C (to reactor)
5) Sludge cooler outlet temperature	(excl. from 1st stage)	25°C	Influent temp. + 15°C
6) Conditioning time in reactor	45 min.	60 - 120 min.	30 - 60 min.

Table 1.2 Main plant facilities

Plant / Facilities & units	Toyohiragawa, Sapporo			Numbu, Fujisawa			Semboku, Sakai		
	Specifications	Q'ty		Specifications	Q'ty		Specifications	Q'ty	
		(a)	(b)		(a)	(b)		(a)	(b)
(1) Appurtenances									
Sludge screen	Automatic bar screen screw press	1	1	-	0	0	-	0	0
Thickner	ø15 m x 3.3 m	4	2	-			ø17.6 m x 5.5 (3.5) m	2	1
Sludge pump	Solid pump, ø80 - 50 x 0.3 m^3/min x 9 m x 3.7 kW	4	2	ø150 mm x 2.5 m^3/min x 17 m x 22 kW	3	2	Solid pump, ø80 - 50 x 0.45 m^3/min x 15 m x 7.5 kW	2	2
No. 1 crusher	Disintegrator, ø200 mm x 30 m^3/hr x 2.4 m x 5.5 kW	3	2	ø200 - ø150 mm x 1 m^3/min x 6 m x 11 kW	2	2	Disintegrator, ø200 mm x 30 m^3/hr x 3.5 m x 11 kW	3	1
Scale preventing device	Electromagnetic type, ø150 x 19 m^3/hr	3	2	-	0	0	Electromagnetic type, ø100 mm x 13.5 m^3/hr	3	1
No. 2 crusher	In-line type, ø200 x 38 m^3/hr x 18.5 kW	3	2	-	0	0	In-line type, ø200 mm x 13.5 m^3/min x 11 kW	3	1

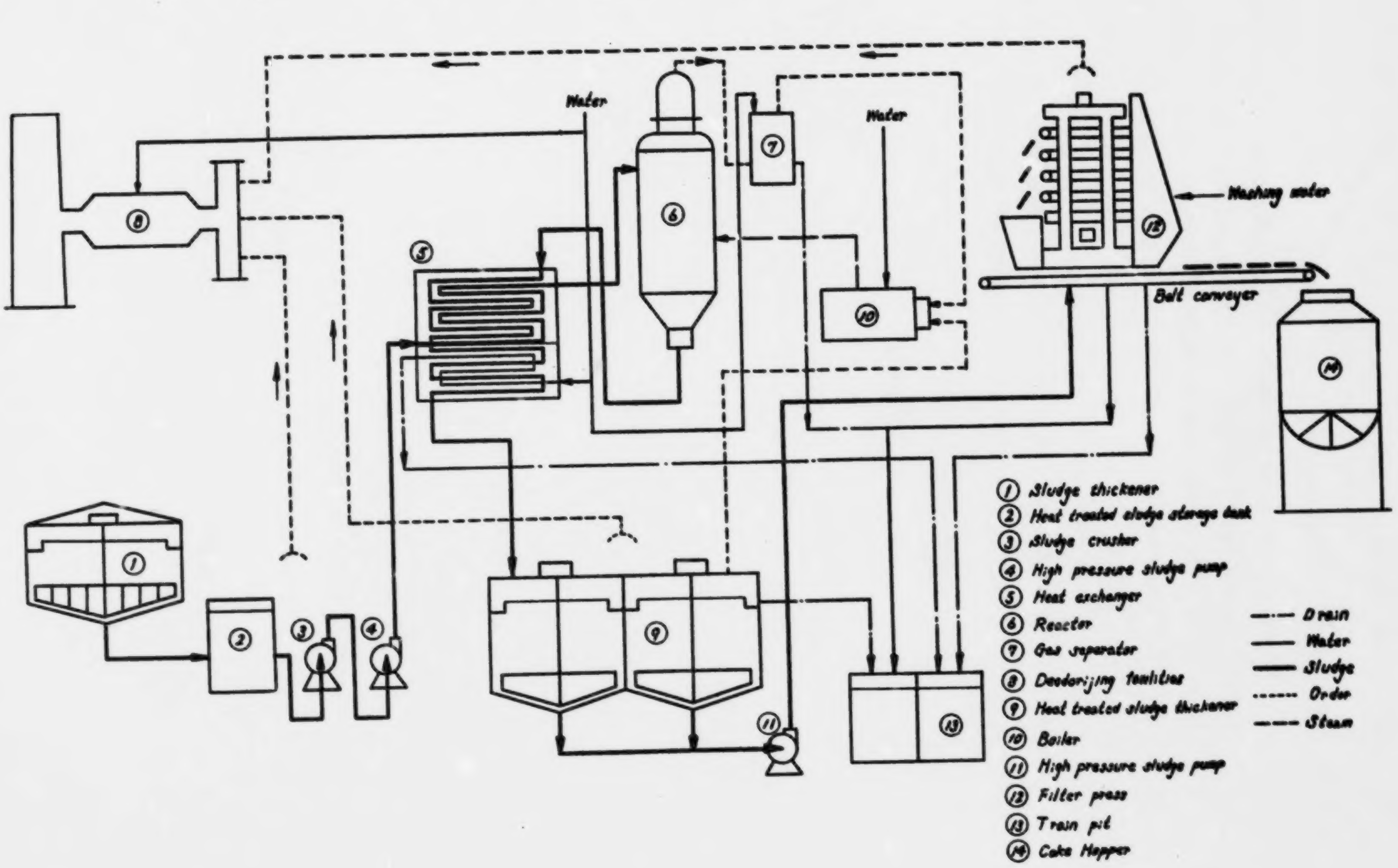
Plant / Facilities & units	Toyohiragawa, Sapporo			Numbu, Fujisawa			Semboku, Sakai		
	Specifications	Q'ty		Specifications	Q'ty		Specifications	Q'ty	
		(a)	(b)		(a)	(b)		(a)	(b)
(2) Heat treatment facilities									
High pressure sludge pump	Diaphragm type, ø65 - ø100 x 19 m^3/hr x 200 m x 30 kW	3	2	Diaphragm type, ø80 mm x 3 - 10 m^3/hr x 30 kg/cm^2 x 18.5 kW	3	2	Diaphragm type, ø65 mm x 10 m^3/hr x 250 m x 15 kW	3	1
Heat exchanger	Counterflow double pipe type, 16.8 m^3/hr x 18 kg/cm^2 x 220 m^2	3	1	Counterflow double pipe type, 3 - 10 m^3/hr x 18 kg/cm^2 x 150 m^2	3	1	Cojnterflow double pipe type, 10 m^3/hr x 18 kg/cm^2 x approx. 165 m^2	3	1
Reactor	Vertical cylinder ø1,798 x 7,000 H x 16.8 m^3	3	1	Vertical cylinder ø1,750 x 7,000 H x 14 m^3	3	1	Vertical cylinder ø1,600 x 7,000 H x 12.5 m^3	3	1
Sludge cooler	Counterflow dual pipe type, 16.8 m^3/hr x 18 kg/cm^2 x 59 m^2	3	0	(Built in the heat exchanger)	3	1	Cojnterflow dual pipe type, 10 m^3/hr x 18 kg/cm^2 x 40 m^2	3	1
Heat treated sludge thickner	ø7 m x 4 m x 154 m^3	2 ss.	1 s.	7,600W x 7,600L x 5,000 W.D.	2 ss.	1 s.	ø10.8 m x 3.6 mH x 325 m^3	1s.	1s.

Plant / Facilities & units	Toyohiragawa, Sapporo			Numbu, Fujisawa			Semboku, Sakai		
	Specifications	Q'ty		Specifications	Q'ty		Specifications	Q'ty	
		(a)	(b)		(a)	(b)		(a)	(b)
Heat treated sludge storage tank	3.5 m x 10 m x 2 mH x 70 m^2	ls.	ls.	-	0	0	ø6 m x 2.7 mH x 85 m^3	ls.	ls.
(3) Sludge dewatering facilities									
Filter press & ancillary equipment	Horizontal type, 1,500 x 1,500 x 50 compartments x filter area 194.5 m^2	5	3	Vertical type, filter area, 25 m^2	2	1	Horizontal type, 1,500 x 1,500 x 75 compartments x filter area, 292 m^2	3	1
(4) Steam generating facilities									
Heavy oil fired boiler		2	2		2	2		2	2

Note: (a): Overall

(b): Number of installations this time

Fig 1-1. Example of process diagrames of sludge treatment section



2. Results of operations and problems

2.1 Results of operations

The three plants surveyed where put in operation on different dates as shown in Table 2.1. Their operational results in one year or two are summarized in Table 2.2, from which it appears that the sedimentability and dewaterability of sludge have been largely improved by heat treatment, and that dewatering by filter press has reduced the moisture content of cake to some 36 to 48%, sharply reducing the cake volume to be handled and making it feasible to carry out landfill with cake or burn it without any fuel.

Table 2.1

Plant	Toyohiragawa, Sapporo	Nambu, Fujisawa	Semboku, Sakai
Construction started	April. 1970	June, 1971	Nov., 1970
Operation started	Mar., 1971	Mar., 1972	Oct., 1971

2.2 Problems on operations

The problems which have been concerned with the heat treatment system are as follows.

2.2.1 Erosion and scale deposition of heat exchanger

At Toyohiragawa Plant, the heat exchanger* in its intial 18 months of operation had part of the extension of the inside tube and around T-tube in the bottom worn out to break part of the latter.

As a temporary repair, a sleeve tube was put on the broken part, along with such measures as improvement of grit removal units, installation of cushion tank, and injection of water at reactor outlet.

At Nambu Plant, after 6 months of operation, some 0.7 mm of abrasion was noticed at two places in the inside tube on the high temperature section, and the tube was renewed accordingly. Also, the sludge crushing pump and heat treated sludge discharge valve were found scored.

At Semboku Plant, after one year of operation, one out of 108 inside tubes in the high temperature section developed an abrasion of some 0.7 mm on the outside, and was renewed. Also, the blades of sludge crusher were found fretted.

The troubles common to all the three plants were baking-up of organic substances to the heat exchanger, which resulted in degradation of heat conduction and increase in heavy oil consumption, not to say deterrence in plant operation.

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* (sludge-to-sludge heat exchanger - hereinafter sometimes referred to as direct type heat exchanger - with raw sludge running through inside tube and heat treated sludge through annular space between the inside tube and outside tube)

Table 2.2 Operational data of heat treatment plants (from April, 1972 to Mar., 1973)

Plant / Item	Toyohiragawa, Sapporo			Nambu, Fujisawa			Semboku, Sakai		
	Max.	Mean	Min.	Max.	Mean	Min.	Max.	Mean	Min.
Average sludge solid, tons/month		562			112			107	
Monthly average moisture content, %		95.3			96.7			96.1	
Heat treatment capacity, m^3/hr		25.0			6.5			10.2	
Heat treating temperature, ^{o}C	200	198	195	201	195	180	200	194	190
Heating time, min.	34	30	27	120	120	60	Apr. – Aug. 29.4 Sept. – Mar. 47.9	27.5 45.5	26.3 42.9
Moisture content of heat treated sludge, %	86.5	84.3	81.4	85.1	78.1	71.3	94.9	87.8	83
Moisture content of sludge cake, %	37.2	36.1	35.1	46.3	37.2	33.1	54.6	47.8	40.6

Plant / Item	Toyohiragawa, Sapporo			Nambu, Fujisawa			Semboku, Sakai		
	Max.	Mean	Min.	Max.	Mean	Min.	Max.	Mean	Min.
Properties of supernatant:-									
Temperature, °C	53	43.3	28	36.1	29.1	22.1	29.0	24.8	19.0
pH	5.7	5.5	5.3	5.8	5.7	5.4	5.8	5.2	4.6
Total solid, mg/lit.	5,950	5,232	4,838	8,100	5,978	4,160	9,899	7,191	4,362
Dissolved matter, mg/lit.	4,040	3,908	3,690	7,900	5,575	3,690	9,252	6,644	3,946
SS, mg/lit.	2,260	1,325	978	650	403	200	1,008	547	118
COD ($KMnO_4$), mg/lit.	1,800	1,590	1,280	4,700	2,975	2,050	3,520	2,615	1,600
BOD_5, mg/lit.	6,000	5,155	3,520	6,100	4,413	3,400	7,660	5,847	4,204
T-N, mg/lit.				1,100	664	410	1,349	704	258
Properties of effluent:-									
BOD_5, mg/lit.	17.3	11.4	5.2	25.0	14.4	5.5	18.2	12.8	7.9
SS, mg/lit.	41.4	25.9	18.3	29	14.6	7	34	19.5	9
Supernatant treatment	Supernatant diluted 300%, aerated 24 hrs, and then fed back to pre-aeration tank			Directly discharged to raw sewage			Directly discharged to raw sewage		

2.2.2 Smells

The sources of offensive odors were waste gases mainly from the reactor and thickener and partly from filter room and cake hopper. Other plant equipment were piped together to form a closed system, and were scarcely any outlet of such stink.

In each plant, the waste gas from the reactor was run through a gas separator and fired in a heavy oil fired boiler or incinerator, together with the gas coming from the thickener. Smelly gas from filter and cake hopper was burnt partly in the incinerator and partly vented out of the stack (Toyohiragawa Plant) or was scrubbed with water and sprayed with deodorant (Nambu Plant).

2.2.3 Treatment of supernatant

Table 2.2 shows the results of quantitative analyses of supernatant and effluent before and after heat treatment at each plant. The volume of the supernatant to be handled was largely dependent upon the heating temperature, heating time, and thickening rate of raw sludge and heat treated sludge, but was about 0.5 per cent of the inflow on the average. The characteristics of supernatant were roughly represented by the following, though different according to specific conditions.

pH : 5 - 6
COD ($KMnO_4$) : 1,300 - 5,000 mg/lit.
Total solid : 4,200 - 10,000 "
BOD : 3,500 - 8,000 "
T - N : 300 - 1,400 "

The supernatant is high in concentration, and if it were returned to the intake of the plant for retreatment together with raw sewage, BOD load would be sent up by 10 to 20%.

At Toyohiragawa Plant, the supernatant was diluted thrice with plant effluent, subjected to 24 hrs of aeration and returned to the preaeration tank, turning out satisfactorily processed effluent.

At Nambu Plant, the supernatant was directly discharged to the intake of the plant, developing some smell and dark brown hue in the effluent.

At Semboku Plant, the supernatant was directly discharged to the raw sewage, and an experiment on the treatment of supernatant from heat treatment process by a step aeration process was carried out by making use of one of the existing tanks with significant results.

2.2.4 Noise

High-pressure sludge pump, air compressor, boiler and sludge discharge valve were noise sources.

The high-pressure sludge pump, for example, roared at a noise level of some 80 phon 1 m apart. But those noise-generating equipment were all hived into an underground cell with its ceiling lined with sound-proof materials, and served no problems to the nearby inhabitants.

2.2.5 Plant maintenance and operation

Toyohiragawa Plant has been maintained and operated by the officials of Sapporo Municipal Government.

Semboku Plant was operated by a private company during tiral run, but, now is operated and maintained by the officials of Sakai Municipal Government.

On the contrary, the operation and maintenance of the Nambu Plant has been consigned to a private business because of difficulties in keeping operators.

Toyohiragawa Plant was stopped for 7 days because of pipe breakdown troubles and another 7 days for two periodic inspections (14 days in total). During the period, sludge was stored.

At Nambu Plant, 10 days were wasted away for reasons of machine troubles, and 17 days were spared for periodic inspection. During the plant suspension, the sludge treatment was taken over by the now stand-by facilities used for chemical dose and vacuum filtration.

At Semboku Plant, 19 days were wasted away for periodic inspection, and sludge during the period was stored as in the case of Toyohiragawa Plant.

Since 10 to 20 days are necessary for periodic inspection and repair, at least one train of standby facilities is indispensable, together with a reservoir to store sludge which will result from the reduction of plant capacity during that period.

Another problem is the compulsory manning requirements. According to the "Pressure Vessel Safety Rules", this kind of heat treatment facilities is required to have a certified chief engineer for boiler operation and a certified chief engineer for danger handling.

3. Tests and investigations for improvement of heat treatment process

In order to settle the problems which were turned up by the running of the full-scale plants, the members of the subcommittee took the lead in conducting some fundamental experiments.

3.1 Wear, corrosion and baking of organic substances in the heat exchanger

3.1.1 Improvement of heat exchanger

The heat exchangers employed were originally of the direct type, as shown in Figs. 3.1 and 3.2, in which the inner tube carried low temperature raw sludge while the annular space, between the inner and outer tubes conveyed high temperature heat treated sludge.

Rather simple in construction though it was, wear, corrosion and baking of organic matter were brought about in the annular space and T-tube which were conveying heat treated sludge.

To solve these problems, the following measures were provided, and the results were analyzed after three months of operation.

(a) Conversion of heat exchanger from direct type to sludge-to-water type (indirect type) as illustrated in Fig. 3.3. With this modification, the heat treated sludge could always be run through the inner tube without necessity of negotiating difficult places where flow pattern was sharply changed in section or direction, thus smoothing the flow and reducing turbulence.

(b) Installation of a cushion tank just upstream of the automatic sludge valve.
With this, abrupt pressure change at the time of sludge valve operation could be abated, minimizing abrasion and corrosion.

(c) Pre-cooling of heat treated sludge by water injection at the outlet of reactor.

(d) Adoption of a special cleaning method in which a cleaning bullet is forced through the inner tube hydraulically with between the bullet loading and unloading ports prepared at the inlet and outlet of the heat exchanger. With this, the cleaning time was saved, doing away with the overhaul of the heat exchanger.

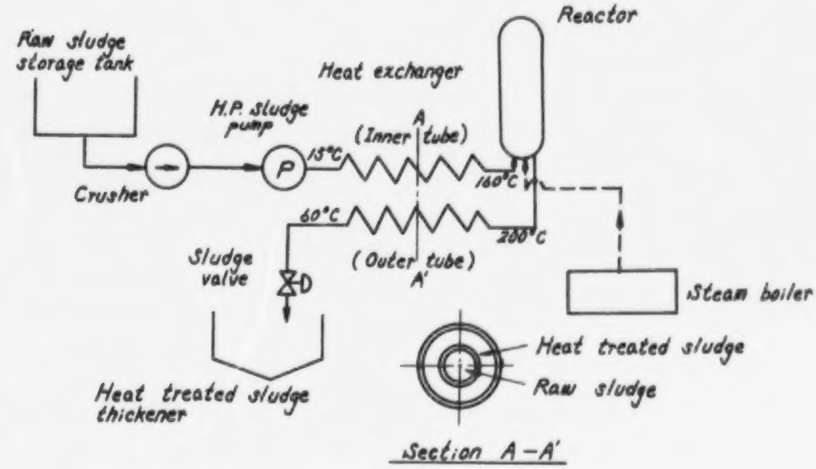


Fig. 3.1 Process flow sheet of direct type heat exchanger

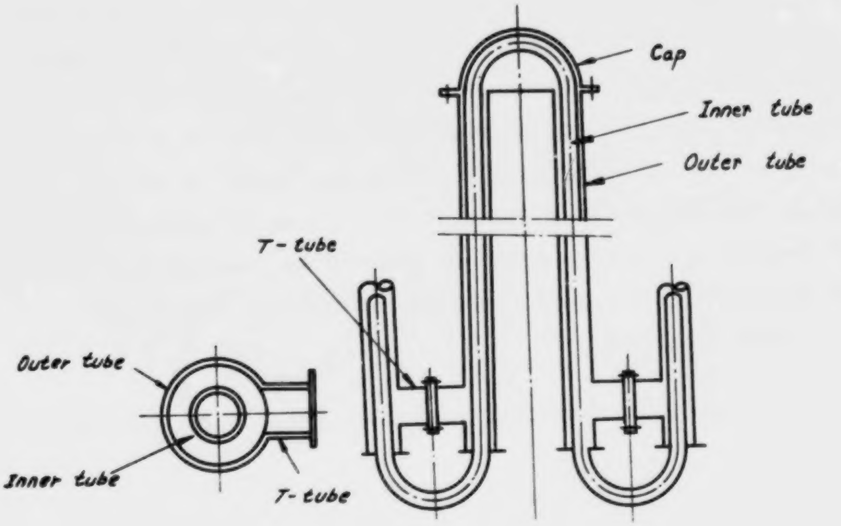


Fig. 3.2 Detail of direct type heat exchanger

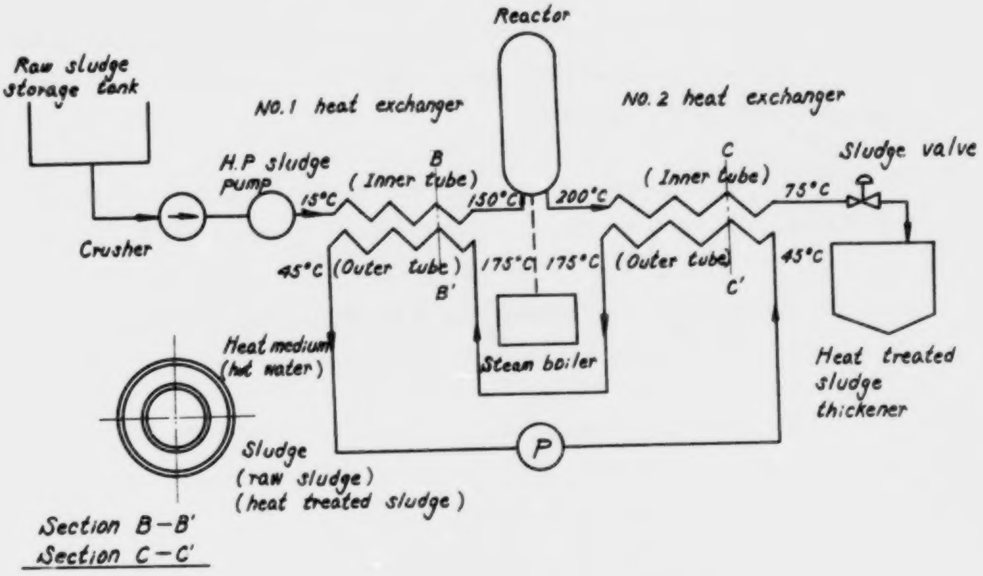


Fig. 3.3 Process flow sheet of indirect type heat exchanger (sludge-to-water)

In three months after improvement, the heat exchanger was disassembled and examined. It was found that the high temperature section of the inner tube was covered with a black hard organic scorch of 0.1 to 0.2 mm in thickness while other parts were laid with the same but 2 to 3 mm in thickness. There was no abrasion nor corrosion to make mention of. The low temperature section had a soft organic layer of 0.1 to 1.0 mm thick, but had neither abrasion nor corrosion whatsoever. These depositions could easily be removed by running the cleaning bullet. The outer tube was covered with an oxide film in good condition.

With all these, the improvement measures taken were verified effective.

3.1.2 Corrosion test within reactor

The heat treating reactor is ill situated so far as corrosion is concerned, because it is always to bear the brunt of high temperature and high pressure in addition to constant attack from chemically active water which contains much solid and is not deoxygenated.

For this reason, the reactor corrosion problems were studied from the metallurgical point of view by conducting corrosion tests on mild steel for boiler use (SB42 - JIS G3103) in reactors at Toyohiragawa and Semboku while stress corrosion tests were conducted at Nambu using stainless steel (SUS32).

At Toyohiragawa and Semboku, mild steel test pieces indentical to reactor material were set in a running reactor for 12 months. The results were as follows.

(a) The test pieces were deprived of rusty layer, and their remaining thicknesses were measured. The measurements proved to be almost the same as before test, evincing that the corrosion rate was very small.

(b) Microscopically, the corrosion was found uniform over the entire surface of each test piece.

(c) Also, it was found that the test pieces were protected with a firm film of Fe_3O_4 to reduce corrosion to a minimum, and therefore that the reactor material was strong enough to sustain the corrosion under service conditions.

At Nambu Plant, three kinds of stainless steel test pieces (SUS32, SUS27 and 18Cr steel) which were prestressed were set in a reactor for 9 months for corrosion test. The results were as follows.

(a) SUS32 which was the same material as reactor's was free from any problematic corrosion symptoms like stress corrosion cracking, pitting and inter-granular corrosion.

(b) SUS27 yielded to stress corrosion cracking, and was judged unfit for the reactor.

(c) 18Cr steel showed no stress corrosion cracking or pitting and were considered to be a substitute for SUS32.

3.2 Deodorization

As already mentioned, each plant was burning away smelly gases of the heat treating process in boiler or incinerator after possing through a gas separator. In order to realize a more effective way of odor removal, catalytic combustion method and ozone oxidation method were studied at Semboku Plant for 2 years. On the other hand, Nambu Plant was applied with a so-called scrubbing method.

3.2.1 Catalytic combustion method

This method was picked up because it was first believed that the plant economy would be improved if smelly gas could be disposed of catalytically at some 200°C as 200°C steam source was already available for the heat treating process.

Odorous gases from reactor, heat treated sludge storage tank, heat treated sludge thickener and supernatant storage tank were forced into a 400 ø x 2,400 mm catalyzer-packed column and burned at 200°C to 300°C. Then, combustion gas was analyzed.

Instinctive test verified the complete removal of odor when the combustion was done at 200°C to 250°C.

The combustion gas analyses disclosed that H_2S and NH_3 were completely removed at 250°C to 300°C and 200°C to 300°C respectively while CO was removed some 80% at 200°C to 300°C and hydrocarbons removed some 50% at 300°C.

A gas chromatographic analysis showed the peak of hydrocarbon spectrum was shifted toward smaller molecular weight since molecular chains were disjoined by catalyzer. Compounds which were identified as responsible for offensive odors included NH_3, H_2S, ethyl amine, ethyl mercaptan, diethyl amine, propyl mercaptan, etc.

NH_2 - group and SH - group of amines and mercaptans were found on a gas chromatograph to have their peak reduced or completely extirpated when passed through the catalyzer layer, verifying that the catalyzer was effective to kill radicals responsible for smells.

The catalytic combustion method was thus justified as an effective way to abate stink generated from the reactor.

3.2.2 Ozone oxidation method

An ozone oxidation method was examined because the exhaust from filter room of which majority was occupied by air diluting offensive odor was considered to jumbonize the catalytic combustion system if it was applied. In a 200 ø x 3,600 mm reactor, smelly compounds and ozone were worked upon each other under humidified conditions, and the results were judged excellent on an instinctive test. There was nc sensible trace of odor when ozone was charged several tens of ppm. Gas analyses, however, revealed that the removal rate of total hydrocarbon was only 10 to 15%. Although ozone was useful to cut NH_2 - or SH - group and abate offensive odor, it would have been not so powerful as to dissociate hydrocarbons.

Anyway, the ozone method manifested itself as practically warrantable for odor-killing.

3.2.3 Scrubbing method

At Nambu Plant, high-concentration foul gases coming from reactor and heat-treated sludge thickener were burned in the form of secondary air, and other subtle room odors were led to a scrubbing tower to wash away their soluble compounds with the effluent from the secondary sedimentation tank (1kg water/kg gas).

After scrubbing, the exhaust gas was almost odorless, substantiating the practicability of this method.

3.3 Studies concerning the treatment of supernatant

As explained in the foregoing, the supernatant produced by the heat treatment system is higher in concentration than that from the ordinary sludge treatment facilities. If it is returned to the raw sewage where it joins influent sewage, it will increase BOD load by 10 to 20%.

In order to mitigate the raw sludge treatment system from BOD overload due to the returning of supernatant, the following experimental treatments of supernatant were conducted at the plants.

(1) Conventional activated sludge process (at Toyohiragawa Plant and Semboku Plant)

(2) Step aeration process (at Semboku Plant)

(3) Extended aeration process (at Semboku and Toyohiragawa Plant)

(4) Aerobic digestion (at Nambu Plant)

3.3.1 Conventional activated sludge process

A combined sewage treatment by the conventional activated sludge process of supernatant and sewage running in at a rate of 1 lit./min. was conducted at a pilot plant of Toyohiragawa Plant. 1 to 2 per cent of supernatant was mixed with the influent sewage, though in the actual plant the supernatant was about 0.5 per cent of influent sewage. This is because the experiment was designed to cover a case where it is required to centrally process various kinds of sludge. The BOD loads were as shown in Table 3.1 below.

Table 3.1 Test conditions (BOD loads)

BOD load	BOD kg/kg MLSS/d	BOD $kg/m^3/d$
Control	0.08 - 0.18	0.29 - 0.47
Case 1 (1%)	0.13 - 0.28	0.46 - 0.68
Case 2 (2%)	0.23	0.65

From the tests, following conclusions were obtained.

(1) 1 per cent addition of supernatant brought about some 0.2 $kg/m^3/d$ increase in BOD load if the conventional activated sludge process is applied.

(2) 1 to 2 percent addition made 0.13 to 0.28 kg/kg MLSS/d of BOD load. In case 1, the retention time in the conventional activated sludge process was more than 5 hrs, and BOD removal was some 95 per cent, showing no significant difference from the control.

(3) COD(Cr) was somewhat higher than the control's, but the removal rate was more than 80 per cent.

(4) Effluent presented light yellowish brown or was evidently colored compared with the control.

(5) By the addition of supernatant, nitrogen compounds were dissolved into effluent with 1 to 2 ppm higher in concentration than the control. But, with the readsorption of metals into activated sludge, the concentration of metals in the effluent was held almost constant.

(6) The growth rate of activated sludge was 1.5 to 2 times as much as the control. In case 1, settling characteristic of sludge was not affected, but in case 2, it was degraded.

(7) The above can be summarized that the treatment of supernatant by the conventional activated sludge process is little or no problem if the adding ratio of supernatant is less than 1 per cent, except that the effluent is tinged with light yellowish brown and that the concentration of nitrogen compounds is increased by 1 to 2 mg/lit.

3.3.2 Step aeration process

At Semboku Plant, the supernatant and sewage had been treated together on the conventional activated sludge process, but effluent BOD had been as high as 20 mg/lit., in addition to the problem that the effluent had been tinged with light yellow.

In order to process the supernatant from the heat treatment process which was high in BOD, the step seration process was considered as activated sludge in its log growth phase was considered preferable for the treatment of supernatant which was high in BOD. One of the existing tanks (effective capacity of 3,080 m^3) was modified for step aeration process, and the sludge ratum rate, feeding points and other factors were changed to obtain the optimum operating conditions by a comparative method.

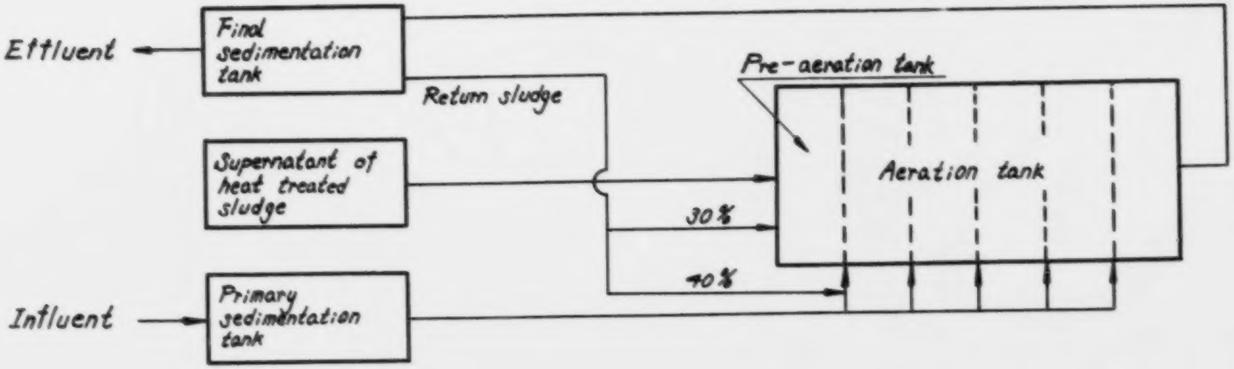
Fig. 3.4 shows a schematic diagram of the step aeration process in which the aeration tank is divided into six sections, the first section of which takes in supernatant and 30% of return sludge, the second section receives the remaining return sludge and one-fifth of the primary and the third through sixth section equally share the remaining sewage among them.

In the first section, supernatant took retention time of 24 hrs, and the combined sewage took 3.4 to 3.9 hrs in other sections. Since SVI was a little increased by the admixture of supernatant, the overflow rate of the final sedimentation tank was reduced to 12.9 $m^3/m^2/d$. Under these operating conditions, the results were acceptable as follows.

(a) Where the supernatant was added 0.52 to 0.72 per cent to the primary effluent, the effluent (from the final sedimentation tank) contained 5.7 to 11.0 mg/lit. of BOD, 5.7 to 18.8 mg/lit of COD ($KMnO_4$) (28.3 mg/lit.COD(Cr)),10 to 15 mg/lit. of NH_3-N, and less than 10 mg/lit. of SS. The effluent was almost colorless and transparent.

(b) Odor from the aeration tank was very little, and the pretreatment tank presented no bubbling.

Fig 3·4 Schematic diagram of step aeration process



3.3.3 Extended aeration process

At Semboku Plant, direct aeration of undiluted supernatant was tried. In an 18-hr aeration, BOD removal was removed 59 to 66 per cent, but violent foaming carried sludge flocs away.

At Toyohiragawa Plant, a spare tank (806 m^3) was used to examine two cases; aeration of undiluted supernatant and aeration of supernatant diluted with the secondary effluent. In undiluted case, BOD was removed about 47 per cent in four days of aeration, and foaming was noticeable just as in the case of Semboku Plant.

When 300% diluted supernatant was aerated for a long time with the aeration set at 3 m^3/hr/aeration tank, m^3, the following results were obtained.

In a retention time of 32.2 hrs, supernatant's BOD was removed 84 per cent, and COD ($KMuO_4$) 28 per cent.

In an additional 24-hr aeration, BOD removal rate reached approximately 90 per cent. But the supernatant, which was high in temperature, generated odorous gas and vapour too much, leaving much to be studied for their removal.

3.3.4 Aerobic digestion process

At Nambu Plant, the supernatant was laboratory tested by an aerobic digestion method.

Fig. 3.5 shows an example of digestion time vs. BOD of mixed liquor and effluent.

The test results were as summarized below.

(a) Aerobic digestion of supernatant under BOD load of 0.1 kg/kg MLSS/d resulted in 95% BOD removal.

(b) Judging from BOD removal and growth rate, the digestion time would require more than 20 days.

(c) Total nitrogen decreased with increase in digestion time; the nitrogen removal was 16 per cent for 20-day digestion and 56 per cent for 60-day digestion.

(d) With increase in digestion time, the hue was improved slightly, but thick brown colour characteristic to the supernatant was still dominant.

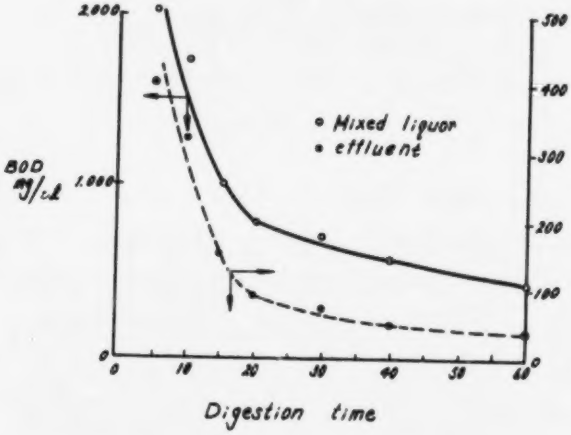


Fig. 3.5 Digestion time vs. BOD

3.4 Studies on the dissolution of heavy metals

At Nambu Plant and Toyohiragawa Plant, studies were made about the effects of heat treatment on the bahaviours of heavy metals contained in the sludge, especially with center around their dissolution into supernatant.

At Nambu Plant, have iron, chromium, copper, cadmium, zinc, lead and arsenic (seven elements in all) were analyzed, and heavy elements - lead, zinc, iron, copper and cadmium at Toyohiragawa.

From the studies, the following were made clear.

(a) The dissolving ratio which is defined as a ratio of the total amount of each heavy metal in the supernatant to that in the raw sludge was 4.19 to 5.41 per cent for Fe, 0.51 to 3.41 per cent for Cr, 4.19 per cent for As, 0.05 to 4.19 per cent for Cu, 0.7 to 9.61 per cent for Cd, 2.67 to 3.70 per cent for Pb, 0.42 to 1.28 per cent for Zn, all in Nambu Plant.
In Toyohiragawa, Fe was 8.00 per cent; Pb, 0.80 per cent; Zn, 1.3 per cent, respectively. Namely, the dissolving ratio of heavy metals and arsenic was less than 10 per cent on the whole.

(b) In spite of heat treatment, more than 90 per cent of sludge heavy metals was retained in the cake to be disposed of. Accordingly, exhaust gas produced by the incineration of cake should be processed through a suitable precipitator in order to trap heavy metals.

(c) Increase in the concentration of heavy metals in the effluent due to dissolution into supernatant registered a maximum of 0.6 mg/lit. for Fe.
Of the heavy metals controlled by the Water Quality Standards for Toxic Substances, such metals as Cr. , Cd and Pb were a maximum of the order of 10^{-4} mg/lit., which was considered not detrimental to the effluent.

4. Cost estimate for installation, operation and maintenance of the heat treating facilities

In order to appraise the economics of the heat treatment process, the capital cost, operation and maintenance costs were estimated for each plant and compared with the anaerobic digestion-chemical coagulation-vacujm filtration process which is now prevailing as a sludge treatment method.

4.1 Method of estimating costs

For each plant, the costs were deflated to those in the 1972 yen, and the costs necessitated for the treatment of supernatant and deodorization were included.

4.2 Capital costs and operation and maintenance expenses

Table 4.1 shows a comparison in capital costs between the heat treatment system and the digestion-chemical coagulation-vacuum filtration system. Table 4.2 shows a comparison between the two systems with reference to operation and maintenance costs and depreciation costs.

Whether the processing is to cover up to dewatering or up to incineration, there is no significant difference in the capital cost per solid tonnage per day between the two systems, except for Toyohiragawa Plant. Up to dewatering, the capital cost reaches ¥ 45 to 47 million, while up to incineration it amounts to ¥ 53 to 60 million.

The costs for operation, maintenance and depreciation are almost the same for both the heat treatment system and the digestion-chemical coagulation-vacuum filtration system if dewatering is included. If extended to incineration, the situation turns out to the advantage of the heat treatment system. But if sophisticated deodorizing facilities, and treatment of supernatant, etc. are taken into account, the advantage cannot always go to the heat treatment system.

In brief, the costs for construction, operation and maintenance are almost the same for both system for the time being.

Table 4.1 Comparison of capital costs between heat treating system and digestion-dewatering system

Plant	Sludge solid, DS-T/day	Heat treatment system			Digestion-chemical coagulation-vacuum filtration		
		up to dewatering ¥ million	up to incineration, ¥ million	Space, m^2	up to dewatering ¥ million	up to incineration, ¥ million	Space, m^2
Toyohiragawa	40.3	(per DS-T) 985 (24.4)	(per DS-T) 1,331 (33.0)	3,150	(per DS-T) 1,390 (34.5)	(per DS-T) 1,801 (44.7)	5,000
Nambu	17.3	820 (47.4)	-	2,460	811 (46.9)	1,039 (60.1)	3,000
Semboku	23.5	1,114 (41.4)	1,234 (52.5)	2,730	1,051 (44.7)	1,329 (56.6)	4,200

B.B.: Values parenthesized refer to unity ton of daily processing sludge (DS).

Table 4.2 Comparison of upkeep costs and depreciation costs between heat treating system and digestion-dewatering system (¥/DS-day)

Plant	Heat treating system						Digestion-dewatering system					
	Up to dewatering			Up to incineration			Up to dewatering			Up to incineration		
	Upkeep	Depr.	Total	Upkeep	Depr.	Total	Upkeep	Depr.	Total	Upkeep	Depr.	Total
Toyohiragawa	6,715	3,620	10,335	7,130	4,950	12,080	7,000	4,070	11,070	9,000	6,160	15,160
Nambu	11,370	6,630	18,000	-	-	-	10,640	5,590	16,230	12,600	8,300	20,900
Semboku	-	-	-	6,860	7,640	14,500	9,140	5,220	14,360	11,100	7,650	18,750

5. Conclusions

The following is the summary of the opinions formed by the committee in regard to the heat treatment process.

> The heat treatment process is still in its developing stage, and has many problems, accordingly. The Committee has obtained the first-hand knowledge about the process based on the three-year investigation program. The heat treating facilities themselves will be little problem if proper measures were taken for the prevention of corrosion and organic deposition. Dewaterability of heat treated sludge is excellent, and dewatered sludge can burn well without any additional fuel. But, the following problems still remain unsettled.
>
> (1) Deodorization of stink from heat treatment process.
>
> (2) Treatment of supernatant with high BOD (incl. removal of heavy metals and nitrogen compounds).
>
> (3) Establishment of operation and maintenance system including periodic inspection.
>
> If there are municipalities which are inclined to construct the heat treatment system, they should carefully examine the necessity of sludge incineration, space availability, prospects of manning for system operation and maintenance, and above all the costs for construction, operation and maintenance.

Before the heat treatment system would establish itself and become accepted widely by sewage plants, it might possibly presuppose the following developments, along with the solution of the problems pointed out by the Committee.

(1) Determination of optimum operating conditions

Quality of sludge varies with plants, and the optimum operating conditions should be determined for each plant by basic experiments.

Considering the quality of supernatant, scorching of organic substances and overall economy, the heat treating temperature and time should be as low and short as possible so long as the aimed dewatering rate can be attained. Namely, the heat treating conditions should be optimized in consideration of the entire sewage treatment process.

(2) Establishment of closed system for heat treatment facilities

The cake obtained from heat treated sludge is low in moisture content and high in calorific value. Namely, it is desirable to recuperate the heat from incineration of cake for the heat treatment, to use smelly gases as incinerator combustion air for complete deodorization and also to reduce the discharges only to ash and combustion exhaust.

In this context, the advent of an incinerator with a heat recuperative boiler compatible to the heat treatment process is strongly hoped for.

(3) Unitization of equipment

The standardization and unitization of equipment should be pushed forward for the purpose of simplifying the facilities, facilitating quality control and saving costs.

In support of the conclusions and opinions formed by the Committee, the Ministry of Construction has agreed in principle to recognize the heat treatment system as eligible for government subsidy programs.

To complete the eligibility, however, it is necessary to develop new equipment meeting the abovementioned requirements.

In the pursuit of this purpose, the Ministry is starting the assessment of the development and improvement of equipment, and the Institute of the National Sewage Works Corp. will play a key role in this.

COMBINED TREATMENT OF MUNICIPAL AND INDUSTRIAL WASTEWATER

presented by
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Japanese Government

§ I. GENERAL

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COMBINED TREATMENT OF MUNICIPAL AND INDUSTRIAL WASTE WATER

§ I. GENERAL

1. Treatment systems of industrial waste water
2. Considerations required in combined treatment
    (1) Technical considerations
    (2) Allotment of expenses
    (3) Administrative measures

With the expansion of today's industrial and economic world which is built on such fundamental materials as metals and petroleums, pollution of public waters by industrial waste water has spread not only over heavily industrialized major cities and their suburbs, but also over farming and fishing areas, posing a serious threat to water source and to human health and life.

Under these circumstances, the importance of public sewerage, the role of which is to treat and dispose of municipal waste water which is discharged continuously, has become increasingly large. To protect public waters and human health and life from pollution, work on public sewerage and industrial waste water must be effectively carried out, but it is important that we make efforts to find possible and effective approaches for pollution control in combination of them.

Possible countermeasures may vary according to the circumstances in which a country or district finds itself, but in our present state, we believe it is urgent that better solutions to processing techniques, expense allotments and administrative systems be sought through exchanging information among those who are confronted by similar problems.

1. Treatment systems of industrial wastewater

The biochemical oxygen demand (BOD) load of industrial wastewater is said to be several times more than that of domestic wastewater, suggesting that the former accounts for a significant proportion of the water pollution problem. Furthermore, industrial wastewater is a major source of contaminants which are threats to human health and life.

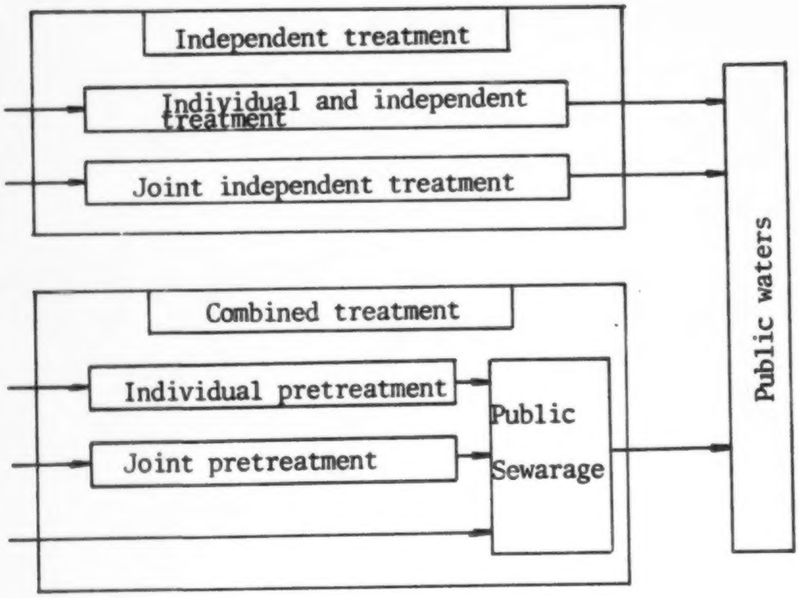
In considering the industrial wastewater problem, possible countermeasures to be taken may be classified as follows:

(1) Independent treatment ... where industrial wastewater is separately treated and then released directly into public waters.
  - a) Individual and independent treatment (A) ... where individual firms and plants treat waste water independently.
  - b) Joint independent treatment (B) ... where wastewater is treated on a joint treatment basis.

(2) Combined treatment ... where industrial wastewater accepted by public sewerage.
  - a) Accepted with pretreatment
    - 1) Individual pretreatment (C) ... where individual firms and plants pretreat wastewater
    - 2) Joint pretreatment (D) ... where pretreatment is made on a joint treatment basis.
  - b) Accepted without pretreatment (E)

The above classification may be illustrated as follows:

Fig. Treatment Systems of Industrial waste water



There are many factors that must be taken into consideration, such as location of factories, and quantities and qualities of sewage water, before selection of a treatment system can be made. To meet requirements of large areas, a combination of two or more of these systems may be employed.

From the standpoint of taking measures based on "control over the sources", independent treatment systems A or B appear to be most desirable, but in many cases, we have to adopt a combination of systems C, D and E.

Where there are no effluent routes other than public sewerage, it may be more efficient in attaining water pollution prevention to take appropriate administrative steps rather than to force small and medium enterprises that can hardly afford, technically and financially, an independent treatment.

We will now discuss some of the problems that may be encountered in employing a combined treatment.

2. Considerations required in combined treatment

The problems in planning of a combined treatment are divided into three principal items; technical measures, allotment of expenses and administrative measures.

(1) Technical measures

It is generally said that "control over the sources of contamination" is the principal rule of combating public nuisances. However, at the same time, a strong case can also be made for the efficiencies of scale.

In air pollution control, no measures can be other than to do something about the very sources of contamination.

This is mainly because collecting emissions, once discharge into the atmosphere, is prohibitively expensive.

In wastewater control, on the other hand, the gravity flow system is a skillful approach to collecting pollutants. The premise of a combined treatment - collecting and combining flows - is physically and economically feasible.

If conditions given below are all met in wastewater treatment, a combined treatment is both useful and practical.

1) Will not damage sewerage facilities.

2) Will not disturb biological treatment process and can meet effluent requirements of treated water.

3) Will not contain dangerous substances in excess of established limits.

In order to satisfy these conditions, the standards of pretreatment techniques play an important role throughout all phases of construction, maintenance and management of a treatment plant.

The following considerations, in particular, are of paramount importance:

(a) Technical and financial guidance and assistance to medium and small enterprises that are producing undesirable industrial wastewater.

(b) Selection and development of sludge treatment and disposal methods employed during pretreatment of undesirable industrial wastewater containing dangerous substances.

(2) Allotment of expenses

Expenses of a combined treatment should be on a "Polluter-Pay-Principle" (P.P.P.) basis. However, because of difficulties in determining quantities and qualities of industrial wastewater running into common public sewerage in early stages of building a treatment plant, the trend in Japan is toward having users share these expenses in the form of fees.

As for the problem of expense allotment, our considerations should not be limited to a simple balance of allotment. A system based on water qualities and a progressive charging system should also be considered so as to provide an incentive for suppressing discharge pollutants.

(3) Administrative measures

For the purpose of increasing effectiveness of a combined treatment, it is necessary that wastewater discharge from individual firms be monitored so that it be kept within appropriate limits, along with enforcement of strict wastewater discharge restrictions including greater penalties to offenders.

For a satisfactory result in administrative measures taken, appropriate guidance and assistance should be employed, rather than relying on restrictions alone. Among such steps may be technical assistance covering improvements in production processes, particularly in medium and small enterprises, and expansion of financing systems for the installation of pretreatment facilities.

It is an effective approach that the municipal authority plans and constructs joint pretreatment systems. And, on a greater scale, basic policies such as formation of new industrial districts in harmony with city planning are necessary.

So far we have discussed basic considerations required for a combined treatment of industrial waste water, and now we will briefly report on some examples of combined treatment in service in Tokyo, Yokohama and Kyoto.

Yokohama Operating conditions of common pretreatment facilities based on the system "D" where wastewater is collected according to the qualities of water to be drained are discussed along with related problems and countermeasures.

Tokyo Chronological processes of common pretreatment facilities that were initially started in the "D" system and were later combined with the "C" system for processing wastewater containing dangerous substances to meet restricted requirements on wastewater discharge, are discussed along with future plans to introduce public sewage treatment plants and the latest technical developments for deep aeration.

Kyoto Results of recent experiments on improving activated sludge methods (double-stage process and oxygen aeration) conducted in the public sewage treatment plant where a combined treatment is being carried out with the systems "C" and "E" for processing industrial wastewater from dyeing plants, are presented.

§ II. TORIHAMA INDUSTRIAL WASTE WATER PRETREATMENT PLANT: HOW IT OPERATES AND PROBLEMS FOR FUTURE IMPROVEMENTS

- The City of Yokohama -

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§ II. TORIHAMA INDUSTRIAL WASTEWATER PRETREATMENT PLANT: HOW IT OPERATES AND PROBLEMS FOR FUTURE IMPROVEMENTS

- The City of Yokohama -

Introduction

At the recent Second U.S.-Japan Conference on Sewage Treatment Technology, combined treatment of industrial and municipal wastewaters in Yokohama City was reported on along with an example of industrial wastewater treatment processes as practiced in Torihama, a coastal industrial district in Yokohama.

Major features and treatment methods reported are summarized as follows:

(1) Features

1) The construction and maintenance costs relating to the joint treatment of industrial wastewater is a full charge to constituent enterprises, and the construction and maintenance activities are placed under the control of the City.

2) Industrial waste water is classified into three types: miscellaneous wastewater (from water closets, kichens, etc.) general process wastewater (containing organic matter, oils, etc.), and pickling and plating process wastewater (discharged from pickling and plating factories). Wastewater from plating and pickling processes is further divided into two types, that containing cyanide and that containing heavy metals.

Each type of wastewater is properly treated according to its particular physical and chemical properties and conditions.

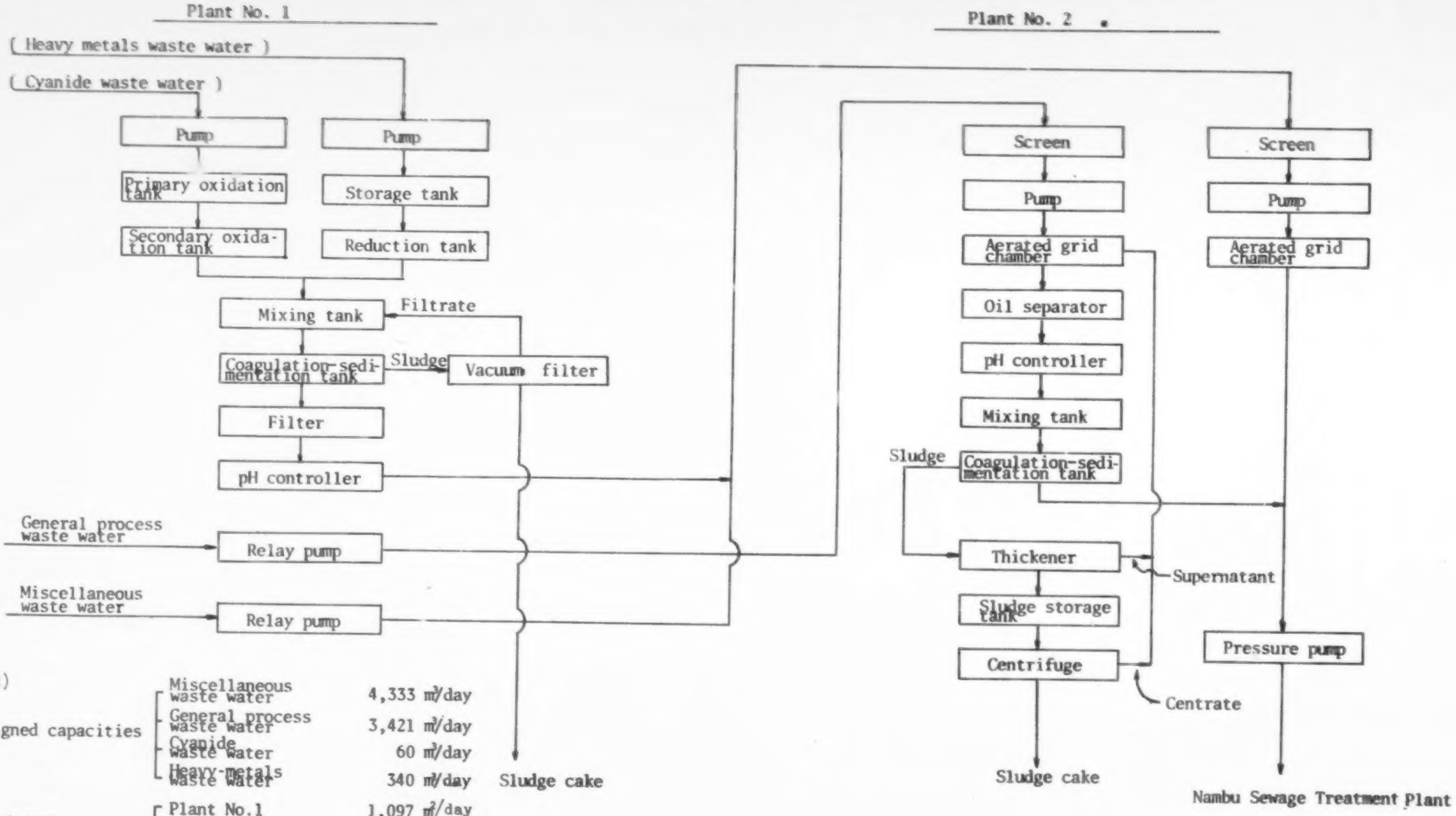
3) Loans of comparatively low interest are being offered by the City to medium and small firms as a public nuisance prevention fund to help them finance their shares of the common wastewater treatment plant construction costs.

4) Miscellaneous wastewater, general process wastewater, cyanide wastewater and heavy-metals wastewater, discharged from the

industrial district, are separately drained through individually provided pipings and then treated separately according to their physical and chemical properties and conditions before they are sent to a sewage treatment plant where all the wastewaters are combined, mixed together and treated by an activated sludge process.

A flow-sheet of these treatment processes is shown in Fig. 1. This paper covers the operating conditions at the Torihama Treatment Plant since the previous report and problems encountered that require further consideration.

Fig. 1 Flow-sheet showing treatment processes at Torihama Industrial Waste Water Pretreatment Plant



(NOTE)

Designed capacities	Miscellaneous waste water	4,333 m^3/day
	General process waste water	3,421 m^3/day
	Cyanide waste water	60 m^3/day
	Heavy-metals waste water	340 m^3/day
Plant area	Plant No.1	1,097 m^2/day
	Plant No.2	3,300 m^2/day

1. Present Servicing Status of Torihama Industrial Wastewater Treatment Plant

The Torihama Treatment Plant No.2 has been in operation since April of 1972. However, due to the fact that the volume of influent has been much smaller than initially expected and that most of the wastewater processed contains oils, the oil separating units have been operated intermittently.

The Torihama Treatment Plant No.1, on the other hand, has been in operation since March of 1973, and following a two month trial period, it has been operating satisfactorily.

(1) The number of constituent enterprises and the amount of wastewater

The number of constituent enterprises as of the end of August, 1973, and the amount of influent classified by it's type are shown in Table 1.

Table 1 Number of enterprises and the amount of wastewater

Types of wastewater \ Items	Enterprises			Wastewater		
	Planned	Existing	Rate(%)	Designed	Present	Rate(%)
Miscellaneous	169	81	48	4,333m^3/day	430m^3/day	10
General process	61	41	67	3,421	586	17
Pickling and plating process	6	4	67	400	270	67
			Total	8,154	1,286	16

Note: Breakdown of pickling/plating process wastewater

Types \ Volume	Designed	Present	Rate(%)
Cyanide	60m^3/day	60m^3/day	100
Heavy metals	340	210	58
Total	400	270	67

As shown in Table 1, the number of constituent enterprises entered has reached approximately 50% of the designed number while the present volume of wastewater comes to only 16% of the designed level. The designed volume has been reached only by wastewater containing cyanide, while at present, the volume of wastewater containing heavy metals is only 58% of the designed volume.

(2) Qualities of the influent and the effluent

The quality of influent and that of effluent at Torihama Industrial Wastewater Treatment Plant are shown in Tables 2 and 3.

Table 2 Record of Treatment of Plant No.1 (Pickling/plating process wastewater)

Item / Date	Type of wastewater	Hue	Odor	Water temp.(°C)	pH	CN (mg/l)	T-Cr (mg/l)	Cr^{+6} (mg/l)	S-Fe (mg/l)	Ni (mg/l)	Cu (mg/l)	Zn (mg/l)	Pb (mg/l)	Cd (mg/l)	Remarks
1973.6.11	Cyanide	Yellow	None	16.0	11.0	190	1.6	-	-	0.9	4.3	110	0.2	0.04	Avg. vol. during June 190 m^3/day
	Metals	Yellow	Cresol	16.5	11.3	13.0	130	-	1.6	20	4.2	17	0.1	Trace	
	Effluent	Light yellow	Slightly cresol	16.5	7.4	12.0	0.4	0	0.1	32	8.5	1.4	0.3	Trace	
6.13	Cyanide	Light yellow	None	19.0	11.1	160	3.6	-	3.6	2.0	6.4	120	0.4	0.02	
	Metals	Yellow	Cresol	19.0	3.2	28.0	51	-	8.2	15	2.9	56	0.1	0.01	
	Effluent	Light yellow	None	19.0	8.2	Trace	1.3	-	0.2	0.5	0.6	.2	0.1	Trace	
6.27	Cyanide	Light yellow	None	21.0	10.9	350	0.7	0	4.0	2.5	3.2	190	0.1	0	
	Metals	Yellow	Cresol	22.0	2.8	8.8	47	22	5.4	14	7.5	43	0.3	0.02	
	Effluent	Light yellow	Slightly crosol	21.0	8.3	4.0	0.2	0		9.4	0.4	0.2	0.1	0.01	
7. 4	Cyanide	Colorless	None	22.0	10.9	-	0.6	-	-	2.9	5.3	120	0.5	0	Avg. vol. during July 240 m^3/day
	Metals	Yellow	Cresol	22.0	2.4	-	37	19	-	16	3.3	55	0.2	0	
	Effluent	Yellow	Slightly cresol	22.5	6.9	0	0.8	0	-	17	5.6	23	0.3	0.03	
7.11	Cyanide	Colorless	None	29.0	10.5	290		-	-	-	-	-	-	-	
	Metals	Yellow	Cresol	26.0	2.3	1.1	36	11	27	15	4.6	61	1.8	0.01	
	Effluent	Light yellow	Mineral oil	28.0	11.7	23.0	0.3	0	0.1	18	4.4	30	0.1	0.02	
7.25	Cyanide	Colorless	None	25.5	11.0	290	0.2	0	2.2	-	5.8	90	0	0	
	Metals	Yellow	Cresol	27.0	6.0	13.0	35	14	3.0	-	38	37	0.1	0	
	Effluent	Light yellow	Slightly cresol	28.0	3.3	0.7		1.8	0.2	-	1.3	0.3	0	Trace	
8. 1	Cyanide	Colorless	None	27.0	11.1	-	0.3	0	2.5	-	3.4	90	0	0	Avg. vol. during August 270 m^3/day
	Metals	Yellow	Cresol	26.0	1.9	-	30	8.5	6.3	-	3.7	50	0.2	Trace	
	Effluent	Light yellow	Slightly cresol	27.5	8.1	5.8	8.4	5.5	0.4	-	4.5	0.7	0	Trace	
8. 8	Cyanide	Colorless	None	27.0	11.0	140	0.2	0	1.8	-	2.3	64	0.1	0.03	
	Metals	Yellow	Cresol	28.0	3.0	0.1	64	52	3.0	-	2.4	32	0.3	Trace	
	Effluent	Light yellow	Slightly cresol	29.0	8.3	0.1	1.4	1.1	-	-	0.1	0.2	0.2	0	
8.22	Cyanide	Colorless	None	29.0	11.6	150	0.5	0	2.8	-	2.1	87	0.1	0	
	Metals	Yellow	Cresol	28.0	2.6	22.0	14	4.8	7.8	-	5.0	22	0.2	0.02	
	Effluent	Colorless	Slightly cresol	29.0	7.4	0.3	0.3	0	0.3	-	4.5	1.0	0	Trace	
9.12	Cyanide	Colorless	None	25.5	9.5	160	0.4	0	1.6	-	1.2	150	0	0.01	Avg. vol during September 270 m^3/day
	Metals	Yellow	Cresol	26.0	3.0	3.5	420	5.5	16	-	15.0	22	0.4	0.01	
	Effluent	Colorless	Slightly cresol	26.0	7.3	Trace	Trace	0	0.3	-	0.3	Trace	0	0	
9.26	Cyanide	Colorless	None	26.0	11.1	190	1.3	0	1.7	-	1.7	110	Trace	0	
	Metals	Yellow	Cresol	26.0	2.7	2.0	120	41	63	-	37.0	80	0.5	0.02	
	Effluent	Light yellow	Slightly cresol	26.0	8.7	0.1	1.5	1.1	0.3	-	Trace	0.4	0	0.01	
	Desired level				5-9	1 and less	2 and less	0.5 and less	10 and less	-	3 and less	5 and less	1 and less	0.1 and less	

NOTE: Waste waters containing cyanide and heavy metals are treated respectively before they are mixed together, and then conveyed to Nambu Treatment Plant.

Table 3 Record of Treatment at Plant No.2 (general process waste water)

Item / Date	Types	Hue	Odor	Water temp.(C)	pH	BOD (mg/l)	COD (mg/l)	SS (mg/l)	I_2 demand (mg/l)	Cl (mg/l)	Oil (mg/l)	Remarks
1973.6.20	Influent	Turbid black	Mineral oil	22.0	9.9	540	180	650	13	4,100	350	Avg. vol. during June 350 m^3/day
	Effluent	Turbid black	Mineral oil	22.0	9.1	300	140	120	15	2,900	100	
7.18	Influent	Turbid grey	Mineral oil	25.0	9.1	170	140	240	67	6,900	170	Avg. vol. during July 450 m^3/day
	Effluent	Color-less	Mineral oil	25.0	8.6	180	160	78	43	4,800	15	
8.15	Influent	Light grey	Mineral oil	29.0	7.7	340	104	1,900	6	6,100	1,200	Avg. vol. during August 590 m^3/day
	Effluent	Dark grey	Mineral oil	26.0	8.5	110	54	280	25	4,300	140	
9.19	Influent	Turbid grey	Mineral oil	24.0	7.6	210	153	270	22	2,600	110	Avg. vol. during Sept. 607 m^3/day
	Effluent	Turbid grey	Mineral oil	24.0	7.5	170	147	28	140	3,900	43	
Desired level					5~9	300 and more		300 and more			35 and more	

As indicated in Table 2, the cyanide content in the treated water exceeded the desired level particularly during the early periods of plant operation. This was due to the fact that cyanide had been discharged and mixed with waste water which contained heavy metals as a result of misoperation by some station operators. After giving them proper instructions, these kinds of accidents have been considerably reduced and are rarely encountered today.

You may notice that the 6-valent chromium content in the treated water exceeded the desired level on the 25th of July, 1st and 8th of August, and 26th of September. This was supposedly due to an excess amount of sodium hypochlorite introduced for decomposition of cyanide, causing the reduced trivalent chromium to be oxidized again to 6-valent chromium when mixed with chromium containing wastewater. Also, unsatisfactory treatment of copper is generally attributable to difficulties in the formation of cupric hydroxide through coagulation as a result of the formation of complex cyanide.

In Table 3, it can be seen that the concentration of oils and fats in treated water exceeds the desired value. The major causes are, it is assumed, (1) degradation of efficiency in removing oils because of processing various kinds of oils flowing into wastewater being processed, (2) shortened retention period due to, a short-circuit produced in the separation tank, and (3) insufficient oil separation in the natural aeration system.

Although removal of B O D and S S is today done only by gravity s sedimentation rather than a chemical coagulation/sedimentation, obtained values of this plant show that they are well within the desired values.

(3) Operation/maintenance cost

Operation and maintenance costs including chemicals, lighting, heating expenses, power, personnel, sludge disposal and pipe cleaning expenses are totally charged to constituent enterprises.

Allotment of these expenses is based upon the quantities and qualities of waste waters discharged from individual firms. Shown in Table 4 is the Operation/maintenance cost per one cubic meter of wastewater, computed from the present conditions of plant operation.

Table 4. Operation/maintenance cost per 1 m^3 of wastewater

Costs / Types	For present	For designed
Miscellaneous	15 yen	10 yen
General process	Average 60 yen	Average 34 yen
Pickling/plating process	*Average 210 yen	Average 136 yen

* Sludge disposal cost are not included.

The operation and maintenance cost for general process and miscellaneous wastewaters runs higher than its designed cost. This is because the volume of wastewater is comparatively small and the fixed cost, such as personnel expenses, is almost constant irrespective of the amount of waste water. As for pickling/plating process wastewater, the amount of chemicals used is twice or thrice the designed amount, thus pushing up the cost to a very high level.

(4) Operation and maintenance system

Operation and maintenance of plant No.1 and No.2 is conducted by two daytime personnel shifts and one night shift, on a four-man shift basis. Tasks performed by the operators consists primarily of machine operation, feeding chemicals into chemical tanks, maintenance and inspection of equipment and instrumentation, keeping daily operation reports, and conducting simple water quality tests.

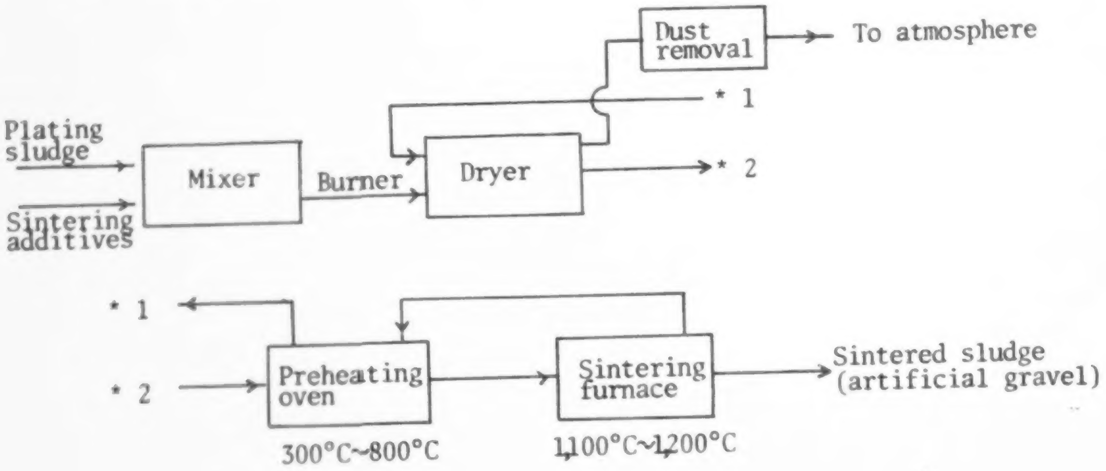
(5) Sludge and waste oils, and their disposal

1) Sludge disposal

During treatment of pickling/plating process wastewater, 600 to 800 kg (water content 80%) of sludge containing such dangerous materials as heavy metals are produced daily. These substances should not ooze out of the sludge when disposed, and we are presently concentrating our efforts on developing desirable methods of disposal.

For the time being, sludge is stored in sealed containers within the sewage treatment plant yard until a desirable method of disposal is found. Under study is a method in which sludge cakes containing dangerous substances are mixed with such additives as clay, glass chips and sludge from waterworks and then heated to 1,100 to 1,200°C in a furnace. During this high temperature treatment, components in the additives (mainly SiO_2, Al_2O_3, Fe_2O_3, etc.) react with heavy metals in the sludge in a solid-state mutual reaction to form a glass-phase, in which heavy metals in the sludge are sealed inside permanently, and the dangerous substances enclosed are thus prevented from oozing out upon disposal. The flow - sheet below shows this process which is now under study.

(a) Sludge treatment flow-sheet under study



(b) Test results obtained

i) The experiment has been conducted by processing dry sludge in a laboratory type fixed furnace.

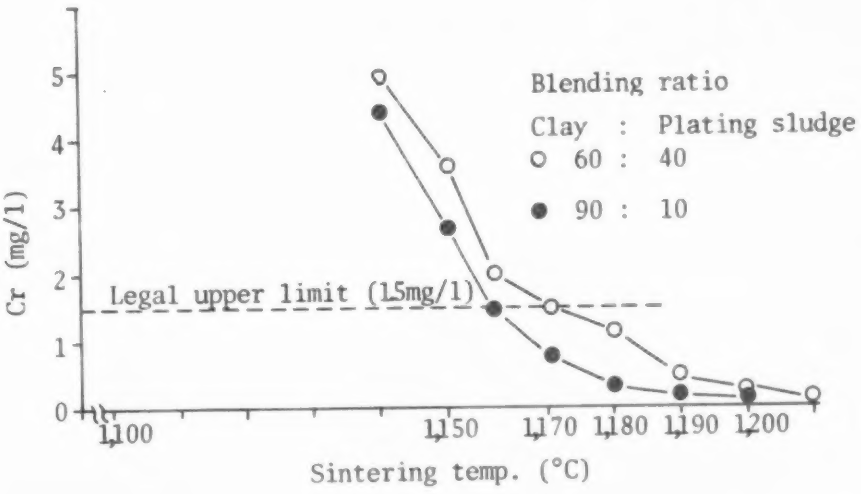
ii) Tabulated below are the sintering additives and temperatures used.

Sintering additives	Sintering temperatures
Clay	1,180 to 1,220
Sludge produced from the municipal waterworks	1,150 to 1,180

NOTE: Blending ratio of plating sludge to sintering additives is roughly 25 to 75

Sintering temperatures decrease with an increase in Ca content, but furnace operation at lower temperatures poses a problem in maintaining desired temperatures because it subsequently reduces allowable baking temperature ranges.

iii) Solubility test of heavy metals from sintered sludge



NOTE: The sample is dipped three hours at 100°C

When sludge is processed at high temperatures, emission of heavy metals into the atmosphere poses a serious problem. We are now studying methods of reducing emissions into the atmosphere to an absolute minimum.

Among the approaches under study are to lower sintering temperatures by adding materials for preventing sublimation of the heavy metals contained, and to sinter heavy metals within a closed circuit furnace so that emission of furnace gases into the atmosphere is totally eliminated.

2) Disposal of waste oils

Waste oils produced during the treatment of general process wastewater amount to 400 to 600 liters a day (water content 50%). These waste oils are removed by waste oil disposal licencees licenced by the Mayor.

In this regard, storage methods of waste oils in treatment plant are subject to the restrictions specified under the Fire Act from the standpoint of fire prevention.

2. Problems Encountered at This Plant and Their Countermeasures

(1) Processing capability of the joint pretreatment plant

The initial cost of the joint pretreatment facilities are totally charged to constituent enterprises and the processing capacity of this plant was decided on the basis of estimates provided in reports from individual enterprises as to the quantities and quanlities of wastewater they were to discharge before the plant was constructed. However, the quantitative and qualitative wastewater data reported were in most cases underestimated, so that today the plant constructed based on these underestimates hardly provides any surplus processing capacity. As a result, it is difficult for the plant to allow constituent enterprises to increace their drainage volumes or change the qualities of their wastewater along with the development and expansion of their business activities.

In view of our experience, therefore, when a new wastewater treatment plant is to be built, it is suggested that the City pay part of the construction cost for the enterprises concerned so that a plant with a capacity flexible and large enough to accommodate future expansion can be built.

At the Torihama Plant, there had been a demand for an increase of 640 cubic meters a day of general process wastewater after the construction of the plant was completed. This demand has been admitted on the condition that additional facilities for this plant will be realized in the near future.

(2) Joint pretreatment of waste water which contains oils

Because a joint pretreatment system has been adopted here for processing wastewater containing oils, the following problems have been encountered:

i) Although oil separation facilities were originally designed as a aeration system, it has been found that oil separation performance is inadequate today, partly because actual influent contains a wide variety of oils, and partly because the concentrations of some of these oils have far exceeded earlier expectations.

To cope with this problem, we plan to take steps which will extend their retention period and to improve plant capabilities upon thorough investigation of physical and chemical properties and inflow rates of oily wastewater which are expected to further increase.

ii) All enterprises which have gasoline filling stations and storage facilities are required under the Fire Act to have oil separation facilities within their premises, but unless management of these facilities is well maintained, there is always a possibility of a large amount of gasoline, oil, etc., in addition to waste-oils from other enterprises, being drained into the sewage pipes.

Since these offenses could lead to fire hazards in sewage pipes and facilities, or present difficulties in proper treatment processes, we have been keeping in touch with enterprise operators and providing them with necessary information by distributing literature and holding guidance sessions.

(3) Facilities for pickling and plating wastewater treatment

1) It was originally intended that pickling and plating wastewater be treated in three separate systems by dividing it into cyanide, chromium and acid/alkaline wastewaters, but due to the complexity and technical difficulties of piping, chromium and acid/alkaline wastewaters are combined and allowed to flow in one piping system. As a result, we face the following problems:

(a) Since heavy-metals wastewater and acid/alkaline wastewater are conveyed in one piping system, trivalent iron ions contained in acid/alkaline wastewater is reduced to bivalent iron ions when reduction of 6-valent chromium takes place. This causes the consumption of reducing agents to increase. Treatment of 6-valent chromium is as follows: it is reduced first by lowering its pH value and then, by increasing its pH value, it is separated and removed as hydroxides. The pH value in acid/alkaline wastewater also changes during this process, so the consumption of a pH adjusting agent increases.

(b) Bivalent iron hydroxide (ferrous hydroxide) is inferior to trivalent iron hydroxide (ferric hydroxide) in coagulation and dehydration properties.

In consideration of the above, it is desirable that pickling and plating process wastewater be processes in three separate systems.

2) Vacuum filters are used for dehydration of sludge, but due to such problems as poor quality filtrate and filtering fabric wash fluids in addition to large volumes of filtrate being

processed, their loads greatly increase when the fluids are returned to the treating system. This makes smooth system operation difficult.

In the meantime, abundant use of calcium hydroxide for increased dehydration efficiency has increased the consumption of sulfuric acid and resulted in production of extra sludge.

We plan to install more dehydrating machines, the number of which will be based on the results of the operation of existing units. Types of new machines will also be thoroughly considered for improvement in overall capabilities at our plant.

3) We require that strong waste liquids produced at plating factories be stored temporarily within the plants and then discharged continuously in small quantities sufficiently diluted with routine wastewater.

In the early days of operations, however, diluting operations by individual firms weren't always performed satisfactorily. As a result, considerable variations occurred and the process efficiencies were affected. Today, however, owing to the intensive guidance we have given operators in factries, this kind of problem is seldom experienced.

In view of the fact that highly concentrated wastewater is small in quantity, it is desirable to treat it separately so that a uniform concentration of sewage can more easily be maintained.

(4) Quantitative and qualitative determination of wastewater for user charges

The volume of drained water at individual enterprises is estimated by means of a water meter installed on the inlet side of its water supply. In cases where users consume clean water in the manufacture of their products, for example, at a raw concrete mill, the quantity of drained water may differ substantially from the supplied water consumption. This is where the difficulty in drained water volume determination lies.

As for the determination of quality of wastewater, individual operators in factories are required to submit a report on it. We found, however, that these stated qualities differ greatly from the fact and lack reliability. Therefore, though time-consuming, it is desirable that determination be made by city officials.

(5) Operation and maintenance costs

Operation and maintenance costs are computed on a liquidation principle at this plant, and this system has experienced such shortcomings as:

1) Operation and maintenance costs vary year by year, therefore their computation is quite complex and time-consuming.

2) Since fixed expenses, such as personnel, lighting, heating and water expenses, remain constant irrespective of the volume of wastewater discharged, early enterprises in the industrial district are charged at a higher rate than new ones.

3) Since the treatment plants are run with expenses totally borne by member enterprises, it is a prerequisite that unanimous consent of all participant enterprises be obtained before giving a go-ahead to any plan for improvement or expansion of facilities.

As a solution to the problems presented above, a new expense allotment system based on revising user charges every two or three years to correct for variations in plant operation expenses may have to be worked out, rather than relying upon a liquidation principle.

(6) Others

Although it is difficult in small scale facilities such as Torihama plant to prepare alternative treatment systems for emergencies, it is desirable to have spare parts for instruments, pumps and valves.

§ III. COMBINED TREATMENT OF INDUSTRIAL AND DOMESTIC WASTEWATER AT SHINGASHI VALLEY

- Tokyo Metropolitan Government -

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§ III. COMBINED TREATMENT OF INDUSTRIAL AND DOMESTIC WASTEWATER AT SHINGASHI VALLEY

- Tokyo Metropolitan Government -

Introduction

Crude trade wastewater discharged from the heavy industrial area in Shingashi valley was a major pollution source to the Sumida River. In order to remove the pollutants and renovate the river, design of a new waste treatment plant was started in 1962. About 730 industries and 200,000 people were to be served by the plant. This was the first attempt in Japan to build a plant for treatment of a mixture of many kinds of industrial wastewater. The plant which is named Ukima has been under operation since 1966. This paper depicts an anecdote of Ukima's rise and fall.

The Ukima Plant was designed to take care of almost all of the industrial wastewater produced in Shingashi valley. Wastes were mainly from metallurgical, chemical, metal plating, food processing, pharmaceutical, pulp and paper, and dye industries. Before the plant began operation the wastes were discharged directly to the Sumida River. At present, the Ukima Plant treats 160,000 cubic meters of liquid wastes daily. Detailed information on design criteria, process basis and historical background of the Ukima Plant was previously reported at the first US - Japan Conference on Sewage Treatment Technology in Tokyo during 1971.

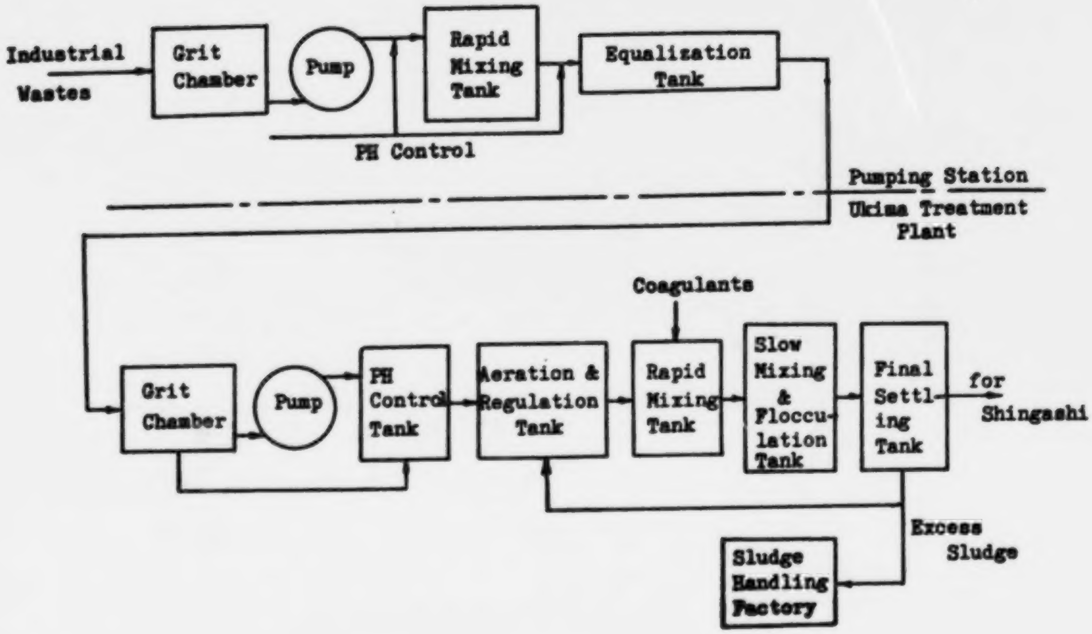

Fig. 1 Flow Diagram of Ukima Plant

Since inauguration of the Ukima Plant operation public concern of pollution has developed very rapidly, with a number of new regulations installed. It may be likened to a "kaleidoscopic change". Living environmental standards, water quality laws, and effluent standard modifications are examples. Pollution is now a popular topic in nearly all daily newspapers and seldom does a day pass without TV or radio reports on pollution. The public is very sensitive to the degree of pollution that exists and this circumstance has resulted in an alteration of the basic idea of Ukima's design basis. Although it consistently serves to save the Sumida River from organic pollution, the Ukima plant is not a panacea. We have come to the conclusion that the (1) effluent standard, (2) sludge handling method, and (3) surcharge system should be revised.

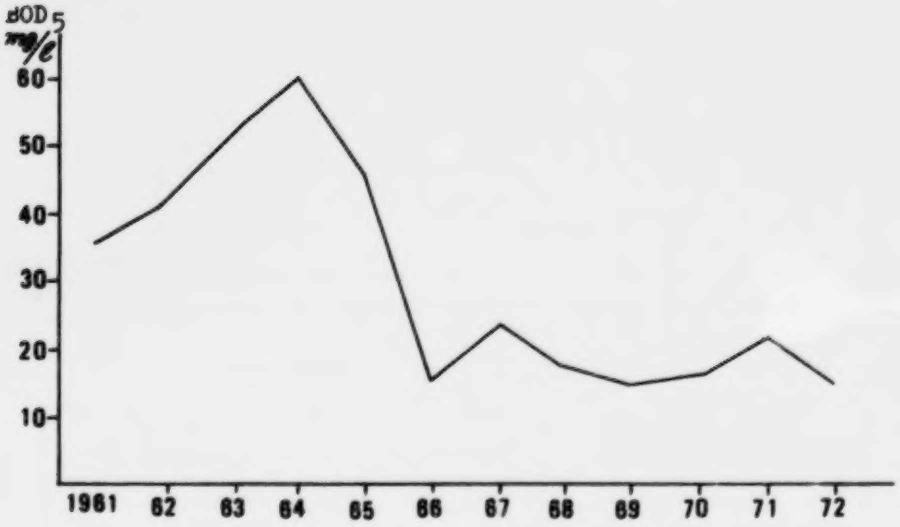

Fig. 2 Stram Quality of the Shingashi River

(the major tributary of the Sumida)

1. Effluent Standard

BOD_5 of 120 mg/l and SS of 150 mg/l were stringent enough for the initial purpose of the Sumida River Pollution Abatement Program. However, in 1971, Tokyo began to believe that the effluent quality should be improved to 20 mg/l BOD_5 and 70 mg/l SS. In addition to these levels the Japanese Central Government regulatory agency announced the following standards to apply to all waste discharges to the River:

(1)	Hexane extract	Mineral oil and grease	5 mg/l
		Vegetable and animal grease	30 mg/l
(2)	Phenol		5 mg/l
(3)	Cyanide		1 mg/l
(4)	Alkyl mercury		N.D.
(5)	Organic phosphorus		1 mg/l
(6)	Cadmium		0.1 mg/l
(7)	Lead		1 mg/l
(8)	Hexavalent chromium		0.5 mg/l

(9)	Arsenic	0.5 mg/1
(10)	Total mercury	N.D.
(11)	Total chromium	1 mg/1
(12)	Copper	3 mg/1
(13)	Zinc	5 mg/1
(14)	Soluble iron	10 mg/1
(15)	Soluble manganese	10 mg/1
(16)	Fluorine	15 mg/1

These are named the "health hazard substances" and the traditional municipal sewage plant is not capable of adequately removing these substances. The only way that the treatment plant effluent quality could satisfy the standard was for the plant to refuse acceptance of the substances. In other words, these materials should be controlled at their source. Every industry is now required to install its own pretreatment equipment for removing the "health hazard substances" before discharging the wastewater to any Tokyo municipal sewer. Table-1 shows the yearly change of the influent quality, which shows a decreasing trend.

Table-1 Yearly Change of the Ukima Influent and Effluent Quality

		1970	1971	1973
PH	inf.	6.9~8.7	6.7~8.6	7.3~8.3
	eff.	7.2~7.6	7.1~7.7	7.4~8.0
SS	inf.	239	203	224
	eff.	71	78	79
BOD	inf.	270	193	197
	eff.	88	64	64
COD	inf.	321	209	198
	eff.	187	97	92
T-N	inf.	85.7	70.5	41.7
	eff.	59.5	57.7	36.9

mg/l

Table-1 Yearly C

Table-1 Yearly Change of the Ukima Influent and Effluent Quality (Continued)

		1970	1971	1973
CN	inf.	0.2	0.2	0.2
	eff.	0.1	0.1	0.1
phenol	inf.	2.3	-	0.4
	eff.	0.5	-	0.1
T-Cr	inf.	4.3	3.6	1.4
	eff.	2.9	1.9	0.9
Cr^{+6}	inf.	-	1.0	0.5
	eff.	-	0.54	0.1
Cu	inf.	2.3	1.9	1.6
	eff.	1.1	0.8	1.4
Cd	inf.	0.21	0.07	0.04
	eff.	0.16	0.03	0.02
As	inf.	ND	0.03	0.01
	eff.	ND	0.03	0.007
Pb	inf.	11.7	6.8	5.5
	eff.	6.3	3.4	2.5
T-Hg	inf.	0.17	ND	ND
	eff.	0.12	ND	ND
T-Fe	inf.	-	-	23.0
	eff.	-	-	16.5
Soluble Fe	inf.	-	-	1.2
	eff.	-	-	2.0
Soluble Mn	inf.	-	-	0.6
	eff.	-	-	0.5
F	inf.	-	-	4.0
	eff.	-	-	4.0
Zn	inf.	-	-	3.8
	eff.	-	-	2.0

mg/l

2. Sludge Handling

Although concentrations of the "health hazard substances" in the influent are less than the standard they accumulate in the primary or excess sludge. Some heavy metals are found at concentrations of 100 to 1000 times more in the sludge as compared to the influent. Analytical data of the heavy metals contained in the sludge are listed in Table-2. Each value in the table is considerably greater than found in normal municipal sewage sludge. Referring to the "industrial and municipal refuse law," one may see that the numbers are so large that the Ukima plant sludge is considered to be "poisonous sludge" which is strictly regulated for final disposal. During ultimate disposal of the sludge special care must be taken to protect surface or ground water from sludge seepage.

Table 2 Heavy Metals Contained in Ukima Sludge and Ash

	date	T-Hg	Cd	T-Cr	Cr^{+6}	Pb	Zn	Cu
Thickened Sludge	71-5	-	516	11,719	-	6,484	-	-
	72-3	-	361	14,562	-	7,522	-	-
	73-6	28.4	150	14,000	0	10,300	2,170	8,950
Ash	72-5	0.024	38.0	6,500	-	15,000	9,500	4,500
	72-10	0.008	56.0	9,000	2,550	11,500	8,000	5,500
	73-6	0.101	47.8	6,580	1,380	4,220	9,490	4,050
	73-6	-	30.8	5,625	1,200	1,875	-	-

mg/kg - dry solid

3. Surcharge

A portion of Ukima's running cost is financed by a surcharge paid by the industries. The basic formula for the surcharge calculation is as follows:

$$C = A + 1.7\,(B + S) + 180P$$

C : Basic factor for surcharge rate
A : Acidity or alkalinity load which exceeds pH 5.6 or 8.7 respectively.

B : BOD (mg/l) fraction which exceeds 300 mg/l

S : SS "

P : CN + Cr, Cyanide (mg/l) fraction which exceeds 2 mg/l
Chromium (mg/l) fraction which exceeds 3 mg/l

In this formula the third item, 180P, on the right hand side amounts to a large portion of surcharge. The original intention of the item was rather for the purpose of rejecting such toxicants as cyanide and chromium, or for applying a penalty for their presence. It was also expected that the penalty would provide a financial incentive to the industries to control strength of their wastewater. However, contrary to the expectation, the industries chose to pay for discharge of the toxicant substances. This resulted in several problems in sewer pipes and plant performance. Therefore, as previously noted, each industry was forced to have its own pretreatment plant and since early 1973 no surcharge has been collected.

However, we are thinking of a new concept of surcharge which is to be applied not only to Shingashi valley but also to the whole of Tokyo. It is based on both BOD_5 and SS concentrations in the wastewater discharge. All waste discharges whose BOD_5 and SS are stronger than the standard domestic sewage are subject to the new surcharge. The new surcharge concept is expected to provide an equitable solution for all users of the Tokyo waste treatment facilities.

4. Shingashi Treatment Plant

When the original Ukima Plant was planned, BOD_5 of 120 mg/l and SS of 150 mg/l were the controlling design criteria of the final effluent. The new requirement for BOD_5 and suspended solids is 20 and 70 mg/l respectively. The Shingashi Plant is now under construction to provide additional treatment of the Ukima Plant effluent. Even though toxic matter is removed at the very beginning, before discharge to the municipal sewer, refractory substances still exist in the influent. With a single-stage biological system it is difficult to produce an effluent containing less that 20 mg/l BOD_5

and 70 mg/l suspended solids. A year of study concluded that if the Ukima plant effluent was diluted with crude domestic sewage it could be made bio-degradable. Thus, a two stage biological system is now intended, consisting of the Ukima Plant and the Shingashi Plant. In 1974, a part of the full system will be operating, and in the near future a tertiary plant may be included.

5. Description of Shingashi Plant

A flow diagram and plant layout are shown in Figure 3 and Figure 4 respectively.

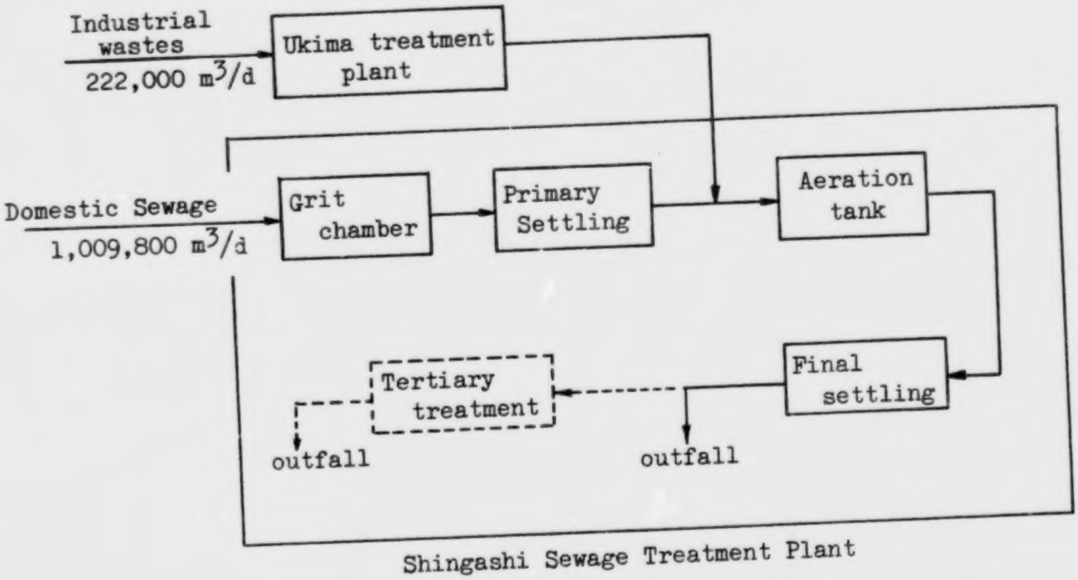

Fig. 3 Flow Diagram of Whole System Plants

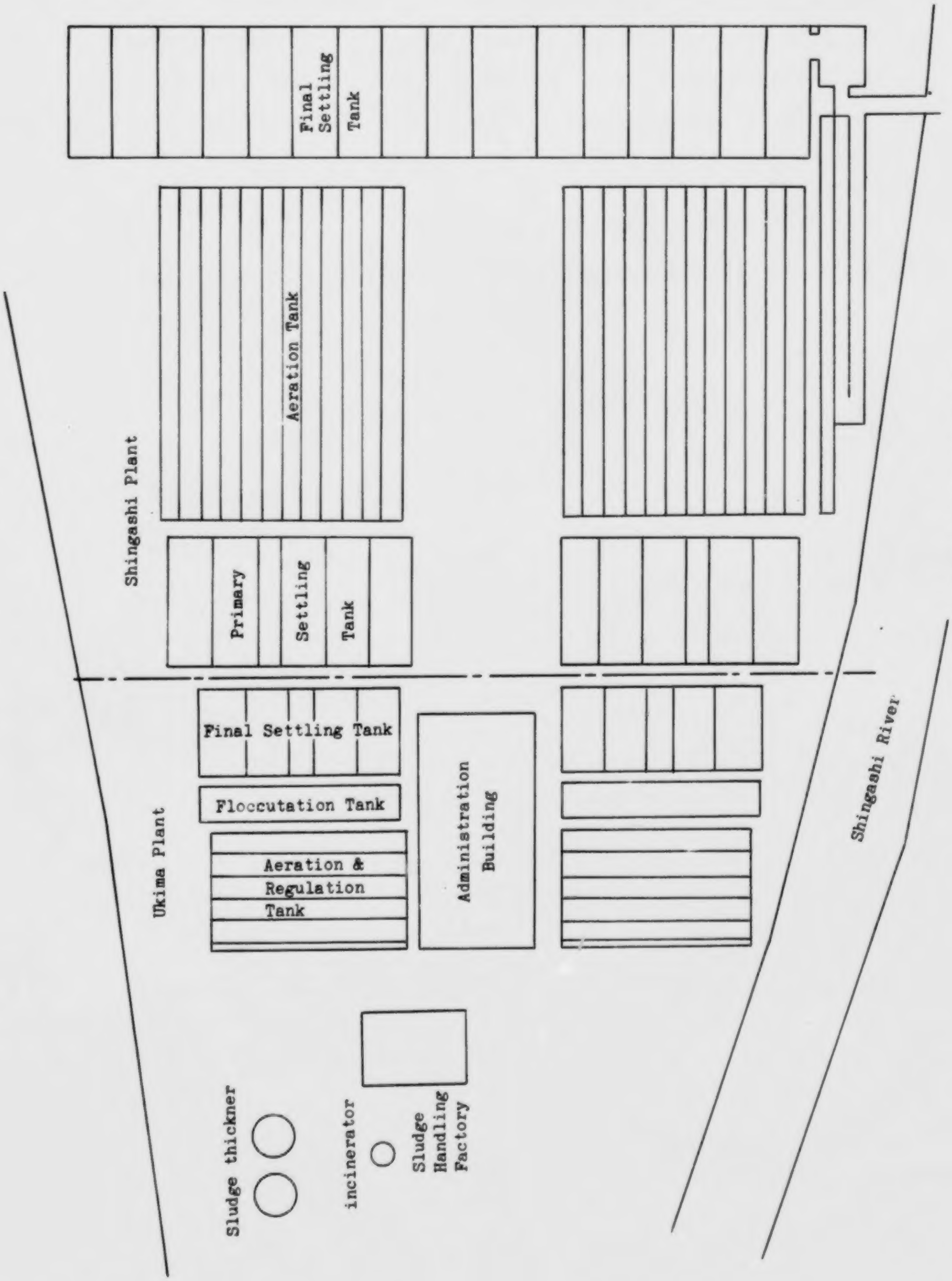

Fig. 4 Ukima & Shingashi Plant

Two level settling tanks (primary and secondary) and deep aeration tanks are interesting design features. The aeration tank is designed to provide a straight one way flow pattern, while traditional aeration tanks have a back and forth flow pattern. Thus a considerable saving of space and head loss are expected.

Table 3 Settling tank in Shingashi Plant

	Primary	Secondary
Structure	2 decks	2 decks
Flow pattern	parallel, one way	parallel, one way
Retention period	1.7 hr	2.5 hr
Overflow rate	50 m^3/m^2/day	34~~50~~ m^3/m^2/day
Dimension, upper	20^W 44^L 3.5^D	20^W 39^L 3.6
lower	20 51 3.5	20 55 3.6

Space saving is one of the important design factors. The local people do not like having a sewage treatment plant for a neighbor. Thus, obtaining land for a sewage treatment plant is a time consuming business. The two deck secondary settling tanks at the Ochiai plant was the first trial of this concept. It was first installed as early as 1962.

For the past three years a deep aeration activated sludge process has been our research subject. Since May, 1973 the pure oxygen process has also been under pilot study. The following comments relate to the development of the deep aeration tank concept.

(1) Oxygen transfer rate, KLa, is proportional to the 0.7 power of the diffuser depth, and electrical consumption is also proportional to the same number. Therefore, energy efficiency is independent of the depth. Table 4 shows changes of KLa relative to the diffuser depth.

The deep aeration system may be considered effective when applied to mixed liquor whose oxygen consumption rate is large. The Shingashi's aeration tank is such a case.

Table 4 Changes of KLa Relative to the Diffuser Depth

Aeration Tank Depth H(M)	Diffuser Depth H'(M)	Tank Volume V(M^3)	Air Volume			KLa	
			Q_1 (NM^3/Hr)	$Q_2=Q_1/V$ (NM^3/Hr/M^3)	Air ratio	KLa(T) (1/Hr)	KLa(20) (1/Hr)
6 (18feet)	5.9	75.4	47.1	0.62	3.1	0.81	0.82
			85.1	1.13	5.6	1.95	1.96
12 (36feet)	11.9	150.7	46.8	0.31	1.6	0.64	0.66
			83.3	0.55	2.8	1.66	1.68
			167.0	1.11	5.5	3.30	3.37
18 (54feet)	17.9	226	60.9	0.27	1.3	1.08	1.09
			122.9	0.54	2.9	5.71	5.78
			227.6	1.01	5.0	9.21	9.28

Air ratio = Ratio of air flow rate to influent flow rate

(2) Aeration tank depths of 60 feet do not significantly influence the activated sludge activity.

(3) When air is diffused at a location deeper than 17 feet, biological floc won't settle in the secondary settling tank due to entrainment of fine air bubbles. Super-saturation causes fine bubbles to accumulate near the water surface and became attached to the flocs. It is the same phenomena that occurs in the air floatation process.

(4) Unless methods of improving the floc settlability obtained, the diffuser should not be submerged deeper than 17 feet; however the tank depth is not limited.

(5) Design criteria for the deep tank, with dimensions up to 30 feet depth, have been derived and confirmed. Use of tanks with depths as much as 60 feet are now being developed.

§ IV. COMBINED TREATMENT OF MUNICIPAL AND INDUSTRIAL WASTEWATER IN KYOTO

- The City of Kyoto -

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§ IV. COMBINED TREATMENT OF MUNICIPAL AND INDUSTRIAL WASTEWATER IN KYOTO

- The City of Kyoto -

I. Present Conditions of the Area Covered by the Sewage Treatment Plant Concerned

I-1. Actual Situations of Factoties and the Quantity of Inflowing Sewage

At the Kisshoin Sewage Treatment Plant, there are two units of the treatment facilities named Route A facilities and Route B facilities, respectively. Route A facilities are for Sujaku Line sewage and a part of Karahashi Line sewage, and Route B facilities are for only Karahashi Line sewage.

Karahashi Line covers an area of 128 hectares in the south of Kyoto City, where are a number of factories for dyeing, electric appliance manufacturing, gilding, and others. As shown by Figure I-1, 48% of Karahashi Line sewage is treated by Route B facilities.

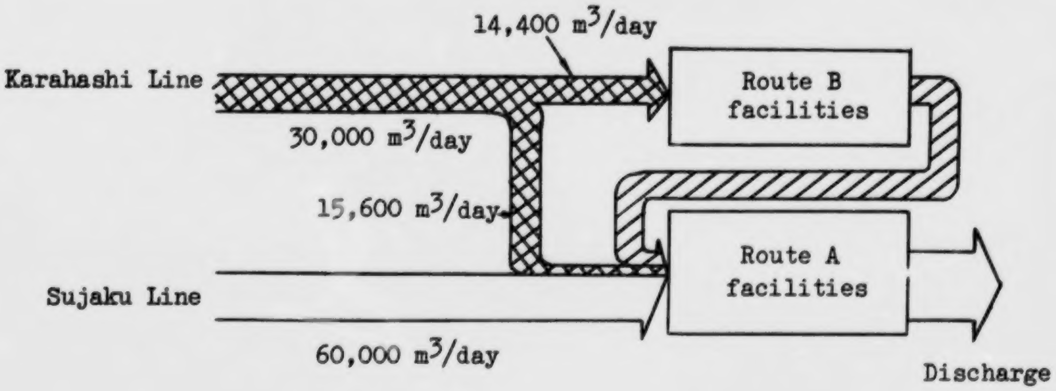


Fig. I-1 Flow Sheet of Kisshoin Treatment Plant

The pH value of the sewage from such factories largely fluctuates and, therefore, the sewage can hardly be treated by the ordinary activated sludge process satisfactorily. As a result, a two stage treatment has been conducted at the Plant now by leading the effluent from Route B facilities to Route A facilities.

Tables I-1 and I-2 show the results of the research which were made in 1969 on the Karahashi Line sewage. According to Table I-1, 41% of the area covered by the Line is residential areas and only 20% is factory sites. Table I-2 indicates that only 13% of the total number of factories is dyeing industry but that it occupies 32%, the largest portion, of the total factory sites.

Table I-1 Area Ratio by Land Use

Factories	Residences	Offices	Roads & Parks	Railway Sites	Total
20.0	41.0	6.0	21.8	11.2	100

Table I-2 Factory Ratio by Industrial Type

	Number of Factories	Area	Sewage Discharged
Dyeing	12.5	31.9	64.6
Machine Mfg.	29.8	31.2	9.1
Metal Work	15.4	3.5	4.7
Food Mfg.	8.1	1.7	10.0
Electrical, Mechanical Work	6.7	3.7	3.3
Others	32.5	28.0	8.3
Total	100	100	100

Out of the inflowing sewage amounting to 30,000 m^3/day, 22,900 m^3/day, 76%, is the waste of factories, while the remaining 24% is the domestic sewage and others. Of the factory sewage, 14,800 m^3/day, 65%, is the dye waste.

I-2. The Characteristics of the Inflowing Sewage

Table I-3 The Quality of the Inflowing Sewage (in 1972)

		Inflow to Route B		
		Average	Maximum	Minimum
Temperature	°C	26.3	30.6	22.0
pH		8.9	9.8	7.4
B O D	mg/l	363.3	632.5	193.8
C O D Mn	mg/l	215.1	348.0	28.1
Total Solids	mg/l	1,461	2,075	834
Fixed Solids	mg/l	1,039	1,287	618
Suspended Solids	mg/l	116	147	77
Soluble Matter	mg/l	1,355	1,875	746
Total Nitrogen	mg/l	38.21	67.51	14.41
Ammonia Nitrogen	mg/l	2.78	9.65	0.16
Albuminoid Nitrogen	mg/l	14.43	45.15	2.20
Nitrite Nitrogen	mg/l	0.27	0.98	0.04
Nitrate Nitrogen	mg/l	0.28	0.56	0.04
Iodine Consumed	mg/l	124.6	300.8	7.9
Chlorine Ion	mg/l	112.9	199.5	12.1
Total Phosphorus	mg/l	2.37	3.00	0.77
Phenol	mg/l	0.35	1.23	0.00
Anionic Detergents	mg/l	8.7	24.0	1.4
No. of Coliform Colonies	colonies/ml	12,000	34,000	19

As shown by Table I-3, the inflow to Route B has been greatly affected by industrial waste, especially by dye waste. Now the special features of the waste quality, which should be taken into consideration for the treatment, and their causes are examined.

(1) pH

Often the inflowing waste is alkaline. Since a great amount of alkali, such as sodium hydroxide (NaOH) and sodium silicate

(Na_2SiO_3), is used for the dyeing, the waste is surmised to be alkalized. At the maximum, pH in the waste reaches 10, but sometimes the waste is acidified to pH 4 to 6.

(2) BOD

Compared with the ordinary municipal sewage, the inflow usually has a higher BOD and its amount largely fluctuates. The average BOD is 300 mg/l, and it often exceeds 400 mg/l. On Sundays and national holidays, it comes down to as low as 100 mg/l. Further even within a single day, it largely fluctuates as shown by Figure I-2. At the peak hour it reaches as high as 600 to 800 mg/l.

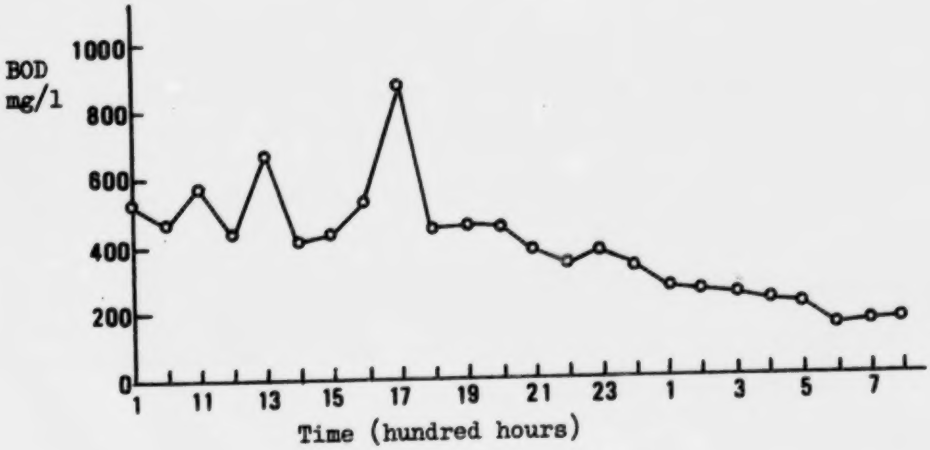


Fig. I-2 Hourly Variation of BOD

In the inflowing waste, 80% of the BOD is soluble BOD. Only 0 to 20% of the BOD can be removed at the primary sedimentation tank and consequently the BOD load in the aeration tank has become high.

(3) Reducing Agent

Sometimes by the waste as much as 300 mg/l of iodine is consumed. It is surmised to be due to the reducing agent contained in the waste. Actually reducing agent, such as sodium thiosulfate ($Na_2S_2O_3$), sodium sulfite (Na_2SO_3), and sodium sulfide (Na_2S), is used as an auxiliary agent for dyeing.

(4) Soluble Matter and Suspended Solids

In the inflowing waste, the total solids is approximately 1500 mg/l on the average. Of the solids, the suspended solids is only about 100 mg/l and most of the solids is soluble matter. Further, of the total solids, the volatile matter is only 35%. It means that the waste is highly contaminated by inorganic soluble matter.

(5) Nitrogen and Phosphorus

There is not much ammonia nitrogen but is a considerable amount of organic nitrogen in the waste.

The amount of phosphorus is a little and the total phosphorus is approximately 2 mg/l.

(6) Coloring by Dyestuff

Since the waste has been colored by dyes, it brings about a feeling of contamination.

II. The Treatment Experiments of Inflow to Route B

Any organisms peculiar to the activated sludge can hardly be found in the aeration tank of Route B. It means that the inflow to Route B can scarcely be treated by the ordinary process. Further if the mixed liquor suspended solid is given more than usual, the activated sludge is decomposed because of poor supply of oxygen, and the treatment efficiency is impaired on a large scale.

At present the BOD removal is 30 to 50% and the SS removal is 10 to 30% by the Route B facilities. Therefore, by existing facilities and at a pilot plant, various experiments were conducted on the treatment of Route B sewage.

II-1. The Treatment Experiment at a Low Load by Existing Facilities

With the quantity of the inflow limited to 200 m^3/hr and with the conditions fixed as shown by Table II-1, an experiment was conducted. The results are shown by Table II-2.

Table II-1 Conditions of Experiment

Inflow	m^3/day	4,260
Air Supplied	m^3 air/m^3 sewage	15.8
Aeration Period	hrs	11.4
M L S S	mg/l	2,336
R S S S	mg/l	6,262
BOD-SS Load	kg/SS kg·day	0.129
Detention Period in Final Sedimentation Tank	hrs	4.10

(1) pH

The pH in the inflow fluctuates between 10.5 and 9.2 and its average is about 10. In the grid chamber, pH is adjusted, but pH in the effluent from the primary sedimentation tank still fluctuates between 8.9 and 6.3. This fluctuation, however, does not hamper the biological treatment. The pH in the effluent from the final sedimentation tank is only 7.5 to 7.1.

Table II-2 Results of Experiment

Hundred Hours	pH			BOD mg/l			SS mg/l		
	Inflow	Primary Effl.	Final Effl.	Inflow	Primary Effl.	Final Effl.	Inflow	Primary Effl.	Final Effl.
9	10.0	6.3	7.4	239	200	19	75	54	12
11	9.9	6.3	7.3	526	283	21	106	146	25
13	9.5	6.7	7.2	206	278	24	73	62	45
15	10.1	8.4	7.1	289	244	24	262	108	47
17	10.4	8.9	7.2	211	241	25	97	84	44
19	9.9	8.9	7.4	252	272	25	81	68	28
21	9.6	8.7	7.5	220	328	23	73	113	23
23	10.2	7.8	7.4	177	223	25	31	22	38
1	10.2	7.9	7.4	206	203	24	20	51	25
3	10.5	8.0	7.4	95	210	21	13	29	50
5	9.2	8.3	7.4	160	142	16	29	53	27
7	10.3	7.6	7.4	137	156	17	18	39	39
Average	10.5~ 9.2	8.9~ 6.3	7.5~ 7.1	226	232	22	74	69	33

(2) BOD

The BOD in the inflow largely fluctuates between 526 and 95 mg/l. The average is 226 mg/l. The fluctuation of BOD in the effluent from the primary sedimentation tank is small being between 328 and 142 mg/l, but BOD has not been removed in the tank. In the effluent from the final sedimentation tank, however, BOD is 22 mg/l on the average having been removed satisfactorily.

(3) SS

Suspended solids in the inflow is between 262 and 13 mg/l and the average is 74 mg/l. SS can hardly be removed in the primary sedimentation tank, and the primary sedimentation tank of Route B functions only to average the inflow quality.

By the conditions that greatly differ from the ordinary activated sludge process, as shown by Table II-1, the ciliata peculiar to the activated sludge are proliferated and the effluent is satisfactory.

II-2. The Two Stage Treatment Experiment by the Activated Sludge Process at a Pilot Plant

The aforementioned experiments prove the fact that to treat the Karahashi Line sewage the load has to be made lower than the standard conditions of the activated sludge process. It is, however, inadvisable to newly build and maintain such low load facilities. Therefore, in the hope that a more effective process might be found, a two stage treatment by activated sludge was experimented.

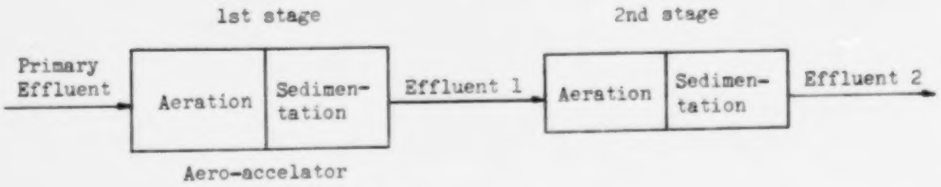


Fig. II-1 Flow Sheet of Two Stage Treatment

As shown by Figure II-1, after the aero-accelator the effluent is to go through another aeration and sedimentation tank. By this method, two experiments were conducted with the operation conditions different from one another.

(1) Experiment I

Table II-3 Conditions of Experiment I

		1st stage	2nd stage
Air Supplied	m^3 air/m^3 sewage	8.2	8.9
Aeration Period	hrs	1.63	4.33
M L S S	mg/l	3,000	1,900
R S S S	mg/l		7,000
BOD-SS Load	kg/SSkg·day	0.89	0.35
Detention Period in Sedimentation Tank	hrs	1.5	1.28

Most of the conditions, as shown in the Table, are standard except the supplied air which is much more than usual. The results are as shown in Table II-4. According to the Table, BOD in the effluent 2 is 18 mg/1 on the average. The BOD removal is 47.4% at the first stage and 75.4% at the second stage. In the second stage there can be seen a considerable number of ciliata peculiar to activated sludge and the result is satisfactory.

Table II-4 Results of Experiment I

	Inflow	Primary	Effl. 1	Effl. 2
pH	11.5 - 4.3 7.5	10.3 - 4.6 7.7	8.8 - 6.9 7.7	7.7 - 7.2 7.5
T S mg/1	725	659	536	482
S S mg/1	68	46	23	21
BOD mg/1	476 - 23 163	324 - 35 139	129 - 14 73	37 - 4 18
BOD Removal %		14.7	47.4	75.4

(2) Experiment II

For Experiment II, five sets of conditions were employed for the second stage while the load of the first stage was kept constant, so that the most effective conditions for the treatment might be found.

Table II-5 Conditions of Experiment II

	1st stage	2nd stage				
	II-1~5	II-1	II-2	II-3	II-4	II-5
M L S S mg/1	3,000	2,000	1,850	1,400	1,200	1,250
BOD-SS Load kg/SSkg·day	0.81	0.150	0.195	0.255	0.302	0.402
Aeration Period hrs	1.0	6.0	5.0	5.0	5.0	4.0
Air Supplied m^3air/m^3sewage	5	14.7	12.2	12.2	12.2	9.8
Det. Per. in Sed Tank hrs	1.15	1.77	1.47	1.47	1.47	1.18
Period of Experiment	Oct 1 ~ Nov 21	Oct 1~ Oct12	Oct 13~ Oct 23	Oct 24~ Nov 2	Nov 3~ Nov 9	Nov 10~ Nov 21

While the experiments were carried out, the quality of each effluent was tested. The results are shown by Table II-6.

The BOD-SS load at the first stage is between 0.78 and 1.13 kg/SSkg·day. The BOD removal at the first stage is approximately 45% at the Experiments II-1 and II-3, but at other experiments it is 20% on the average. At the second stage, the Experiment II-1 shows the maximum BOD removal, 80.9%. At the other experiments it is 68% on the average.

No ciliata peculiar to activated sludge can be seen in the effluent from the first stage, but after the second stage the ciliata can be seen.

The conditions and the results of the two stage treatment experiments at a pilot plant are as follows. At the first stage under the following conditions, the BOD removal is 40 to 50%.

BOD-SS load	approx. 0.8 kg/SSkg·day
Aeration period	1.6 hours
pH	lower than 10

After the second stage, which is operated with the following conditions, the BOD removal becomes satisfactory with BOD in the Effluent 2 being less than 20 mg/l.

BOD-SS load .... 0.15 to 0.2 kg/SSkg·day
Aeration period .... approx. 6 hours
Ca·T (MLSS (mg/l) x Aeration period ... 12,000 – 13,000 up
Air supolied .... 15 m^3 air/m^3 sewage

Table II-6 Results of Experiment II

	II-1	II-2	II-3	II-4	II-5
Dates of Tests	Oct 11 Oct 12	Oct 22 Oct 23	Nov 1 Nov 2	Nov 8 Nov 9	Nov 20 Nov 21
Primary Effluent B O D mg/l	153.3	158.8	186.7	188.1	217.5
pH	10.6-6.0	10.8-5.7	10.7-5.9	11.3-5.1	11.1-4.7
T S mg/l	710	1,032	1,022	1,091	1,167
S S mg/l	84	134	123	130	130
1st Stage B O D mg/l	80.8	124.0	105.4	136.6	195.5
BOD Removal %	47.2	21.9	43.5	27.3	10.2
pH	9.5-6.9	10.2-7.2	9.3-7.0	10.4-7.2	10.3-6.3
T S mg/l	634	964	961	1,009	1,119
S S mg/l	59	96	122	112	127
M L S S mg/l	3,127	3,455	3,580	4,169	3,334
Aeration Period hrs	1.0	1.0	1.0	1.0	1.0
S V I mg/l	63.3	58.8	56.5	53.4	49.5
BOD-SS Load kg/SSkg·day	0.85	0.80	0.90	0.78	1.13
2nd Stage B O D mg/l	15.5	34.0	33.6	41.4	77.7
BOD Removal %	80.9	72.6	68.1	69.7	60.3
pH	7.9-7.6	7.9-7.7	7.9-7.4	9.2-7.9	8.4-7.2
T S mg/l	610	900	913	933	979
S S mg/l	43	68	62	118	86
M L S S mg/l	2,012	1,700	1,372	1,357	2,123
Ca·T	12,070	8,500	6,860	6,790	10,620
Aeration Period hrs	6.0	5.0	5.0	5.0	5.0
S V I mg/l	80.9	75.4	73.1	79.5	73.5
BOD-SS Load kg/SSkg·day	0.123	0.270	0.285	0.373	0.343

II-3. The Treatment Experiments by the Activated Sludge Process with the Pure Oxygen Aeration

As mentioned above, to treat the sewage of Route B by the activated sludge process, a remarkably large amount of oxygen is required in the aeration tank. That is, to supply oxygen enough for the treatment, the amount of air and the aeration period has to be made three times than usual, respectively. It is inadvisable to build and maintain such air system facilities as to meet the requirements aforementioned. Therefore, the pure oxygen aeration process, which was surmised to easily increase the supply of oxygen, was experimented at a pilot plant, which is illustrated by Figure II-2.

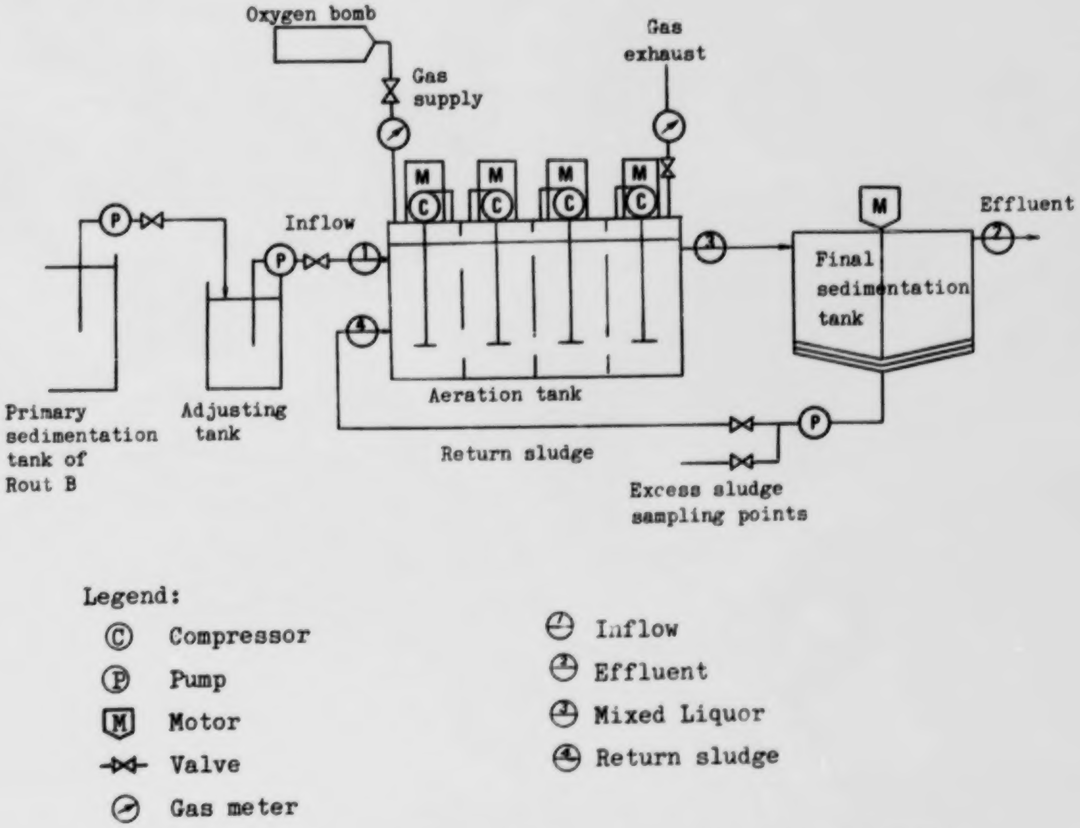


Fig. II-2 Flow Sheet of Pure Oxygen Aeration Process

(1) Conditions of Experiments

The outline of the experiments of the treatment by the pure oxygen aeration process is shown by Table II-7. For the Experiments I and II, the effluent from the primary sedimentation tank of Route B was used as the inflow.

Table II-7 Conditions of Experiments

		Remarks	Period	Quantity of Sewage m^3/d	Aeration Period		Detention Period at Final Sed. Tank hrs
					Q hrs	Q + R hrs	
Acc.		Acclimation of Activated Sludge	31 days Jan 18 to Feb 17	48	3	2.2	3.3
Exp I	1	Quantity of Inflow Constant	16 days Feb 18 to Mar 5	36	4	2.9	4.3
	2		30 days Mar 6 to Apr 4	48	3	2.1	3.2
Exp II	1	Quantity of Inflow Variable	32 days Apr 5 to May 6	48	3	2.2	3.3
	2		21 days May 7 to May 27	37.6	3.8	3.0	4.5
Exp III	A.	Mixed Ordinary Municipal Sewage Inflow Variable	9 days May 28 to Jun 5	48	3	2.4	3.6
	1		9 days Jun 6 to Jun 14	57.8	2.5	1.9	2.9

Legend: Q ... Quantity of inflow
R ... Quantity of return sludge
Acc. & A. ... Acclimation

Firstly, by the Experiment I, the conditions that make the treatment effective at a constant inflow were found out. Then at Experiment II, it was aimed to find out, with the finding of Experiment I made use of, the most appropriate conditions for the treatment of variable inflow which is similar to that of Route B.

The Experiment III was made to see the efficiency of the mixed treatment of Route B sewage and the ordinary municipal sewage. The mixing of those two types of sewage was done in the ratio of 1 to 1 and the hourly variation of inflow was made in the same pattern as that of Experiment II.

(2) The Results of Experiment

To complete the experiments, it took nearly five months from January 28 to June 14 of 1973. Tables II-8, II-9, and II-10 show the average values at each experiment of the operation conditions, the water quality examinations, and the activated sludge examinations, respectively.

Table II-8 Operation Conditions

				Exp. I		Exp. II		Exp. III	
			Acc.	1	2	1	2	A.	1
Quantity of Inflow		m^3/day	48	36	48	48	37.6	47.9	57.8
Return Sludge		m^3/day	19	14	20	17	11	12	17
Ratio of Ret Sludge		%	40	38	43	35	29	26	30
Aeration Period	Q	hrs	3.0	4.0	3.0	3.0	3.8	3.0	2.5
	Q + R	hrs	2.2	2.9	2.1	2.2	3.0	3.4	1.9
Oxygen	Supplied	m^3/day	8.44	8.47	14.00	16.33	16.13	14.67	16.46
	Exhausted	m^3/day	1.61	1.89	2.40	5.89	8.09	8.00	8.85
	Ratio in Exhausted	%	54	56	57	53	56	55	57
	Used	kg/day	10.83	10.60	17.99	17.59	14.15	14.71	16.34
	Ratio of Used	%	89.7	87.5	89.4	78.1	68.3	70.1	69.4
BOD-SS Load		kg/SSkg·day	0.33	0.27	0.37	0.28	0.25	0.34	0.32
BOD-VSS Load		kg/VSSkg·day	0.53	0.41	0.56	0.41	0.37	0.48	0.45
BOD Load Rate		kg/m^3·day	1.67	1.52	2.35	1.80	1.62	1.94	1.87
Final Sedi. Tank	Overflow Rate	m^3/m^2·day	14.8	11.0	15.0	15.0	12.0	15.0	18.0
	Solid Load	kg/m^2·day	103.4	85.3	137.4	131.1	96.8	107.5	135.3
	Detent. Per.	hrs	3.3	4.3	3.2	3.3	4.5	3.6	2.9
	Depth of Sludge Blanket	cm	198	181	175	136	81	110	170
Excess Sludge		m^3/day	0.14	0.15	0.22	0.21	0.07	0.09	0.18
Excess Sludge	SS	kg/day	2.45	3.12	4.65	5.37	2.47	3.26	5.04
	VSS	kg/day	1.77	2.15	3.03	4.07	1.67	2.30	3.61
Exc SS per Rem BOD		kg/kg	0.28	0.38	0.39	0.54	0.26	0.29	0.47
Oxy used p Rem BOD		kg/kg	1.25	1.28	1.42	1.95	1.76	1.32	1.78

Legend:

Acc. and A. .... Acclimation

Exc SS per Rem BOD ...... Excess SS per Removed BOD

Oxy used p Rem BOD ...... Oxygen Used per Removed BOD

Table II-9 Results of Water Quality Examination

			Acc.	I-1	I-2	II-1	II-2	III-A	III-1
Temperature ^{o}C	I		20.0	20.4	21.7	22.1	23.5	22.3	23.0
	E		19.0	19.6	20.8	23.0	24.7	24.1	23.9
Transparency cm	I		3.5	3.0	4.2	5.5	5.3	6.0	5.8
	E		6.5	6.0	9.9	12.6	11.6	18.3	21.6
pH	I	Max	9.0	9.3	10.2	9.3	9.4	8.9	8.4
		Min	3.2	3.6	4.0	5.2	5.7	6.8	7.1
		Ave	7.0	7.6	8.2	7.6	7.8	7.4	7.6
	E	Max	6.9	6.7	8.4	6.7	6.5	6.6	6.7
		Min	5.7	6.1	6.0	6.0	6.1	6.1	6.2
		Ave	6.2	6.4	6.5	6.3	6.4	6.3	6.4
Total Solids mg/l	I		995	1,330	1,340	1,177	850	805	-
	E		790	1,230	1,170	973	780	625	-
Soluble Matter mg/l	I		890	1,180	1,230	1,077	780	750	-
	E		745	1,160	1,120	923	760	605	-
Suspended Solids mg/l	I		82	122	127	92	97	58	75
	E		26	51	49	36	45	19	12
	Rem	%	68.3	58.2	60.5	58.4	53.3	67.5	83.6
Volatile Matter mg/l	I		49	61	65	52	58	43	51
	E		16	23	21	18	22	12	7
	Rem	%	67.4	62.3	65.5	63.0	61.7	72.1	86.3
Dissolved Oxygen mg/l	I		0.0	0.0	0.0	0.0	0.0	0.0	0.0
	E		0.4	0.8	0.7	0.4	1.2	1.5	1.0
BOD mg/l	I		210.5	253.9	328.1	224.8	270.6	242.6	194.4
	E		30.0	23.6	28.0	26.7	13.1	10.7	7.6
	Rem	%	85.8	90.7	92.9	89.3	95.2	95.6	96.1
Soluble B O D mg/l	I		170.4	210.4	255.8	192.3	213.8	199.7	172.5
	E		22.2	11.8	13.7	26.7	9.6	6.7	6.8
COD_{Cr} mg/l	I		432.4	413.0	478.9	521.7	535.3	368.3	364.8
	E		171.0	161.7	139.6	161.6	159.3	99.3	95.5
	Rem	%	60.5	60.9	70.6	67.2	70.2	73.0	73.8
Soluble COD_{Cr} mg/l	I		330.5	277.5	385.8	370.5	416.0	360.0	267.0
	E		152.8	142.5	125.3	120.8	128.3	97.0	86.5
	Rem	%	53.8	48.7	67.3	56.2	69.2	73.1	67.6
COD_{Mn}	I		115.8	125.8	142.6	126.1	131.9	92.0	102.3
	E		65.0	63.0	61.6	65.4	66.1	43.3	45.3
	Rem	%	41.6	49.1	53.2	48.1	50.4	52.8	55.8
Soluble COD_{Mn} mg/l	I		108.8	100.0	104.8	92.8	104.7	84.0	73.0
	E		61.3	50.0	55.5	52.8	53.7	42.0	33.0
Alkalinity mg/l	I		102.5	97.0	187.0	234.0	-	13.0	-
	E		99.0	110.0	217.0	284.0	-	21.0	-
Total Nitrogen mg/l	I		51.27	39.50	51.53	55.90	54.04	61.27	32.34
	E		33.31	33.55	40.65	43.33	53.65	58.63	25.59

		Acc.	I-1	I-2	II-1	II-2	III-A	III-1
Ammonia Nitrogen mg/l	I	2.90	3.13	12.31	8.16	10.16	10.87	6.44
	E	18.86	22.17	25.43	26.11	42.45	42.40	16.64
Albuminoid Nitrogen mg/l	I	39.29	16.76	19.05	33.70	22.51	50.37	25.90
	E	3.19	5.30	3.29	3.67	2.60	16.23	8.95
Nitrite Nitrogen mg/l	I	0.55	0.79	0.93	1.08	0.27	0.41	0.11
	E	0.62	0.62	0.67	0.93	0.30	0.59	0.08
Nitrate Nitrogen mg/l	I	0.05	0.09	0.04	0.18	0.07	0.04	0.02
	E	0.04	0.06	0.04	0.14	0.07	0.06	0.02
Total Phosphorus mg/l	I	2.37	2.71	7.56	4.67	2.92	3.28	2.54
	E	1.03	1.31	6.24	3.83	2.29	2.51	1.64
No. of Coliform Colonies colonies/ml	I	104,000	51,000	27,000	19,800	25,000	55,000	160,000
	E	11,000	6,700	1,800	1,900	840	3,200	3,300

Legend:

Acc. ..... Acclimation
A .... Acclimation
I .... Inflow
E .... Effluent
Rem % .... Removal %

Table II-10 Results of Activated Sludge Examination

			Exp I		Exp II		Exp III	
		Acc.	1	2	1	2	A	1
Suspended Solids mg/l	ML	5,156	5,539	6,472	6,480	6,437	5,739	5,784
	RS	17,855	20,971	22,110	28,417	35,299	35,866	28,012
Volatile Matter mg/l	ML	3,668	3,692	4,210	4,339	4,348	4,054	4,136
	RS	12,846	14,307	14,414	19,061	23,813	25,272	20,047
Ratio VSS/SS	ML	73.5	66.7	65.1	67.2	67.5	70.6	71.5
SV %	ML	45	38	35	29	25	21	23
	RS	96	97	95	96	98	97	97
S V I	ML	91	69	54	44	38	37	39
Sludge Age day	ML	7.7	7.6	7.0	8.6	10.6	12.4	8.0
Dissolved Oxygen mg/l	AT 1	6.0	6.0	6.0	5.3	5.5	5.8	5.7
	AT 2	7.9	7.2	6.6	6.0	6.1	6.4	6.3
	AT 3	9.0	7.8	8.4	7.2	7.8	7.9	8.2
	AT 4	9.4	8.3	8.6	7.8	9.2	9.0	9.3
	A AT	8.1	7.3	7.3	6.6	7.2	7.3	7.4
Oxygen Uptake Rate mg/l·hr	AT 1	23	65	46	62	53	51	51
	AT 2	29	49	49	60	52	60	52
	AT 3	40	39	48	62	51	48	50
	AT 4	31	31	41	50	44	41	48
Sludge Precip. Rate cm/min	ML	1.00	1.48	1.80	2.98	3.53	4.11	3.59

Legend:

Acc. & A	Acclimation
ML	Mixed Liquor
RS	Return Sludge
SV	Volume of Sludge Settled in 30 Min.
AT	Aeration Tank
A AT	Average in Aeration Tank

(1) BOD

As shown by Figures II-3 and II-4, if only the Route B sewage is aerated for 3 hours and its BOD-SS load is 0.3 kg/SSkg·day or over, the BOD removal largely fluctuates and BOD in the effluent exceeds 20 mg/l. If it is aerated for 3.8 hours and its BOD-SS load is 0.5 kg/SSkg·day, the result is satisfactory as its BOD removal is 90% up and BOD in the effluent is less than 20 mg/l.

Fig. II-3 Relation between BOD-SS Load and BOD Removal

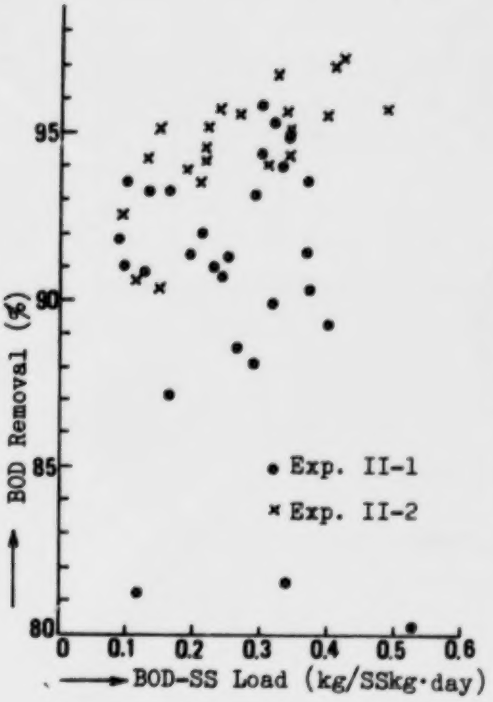
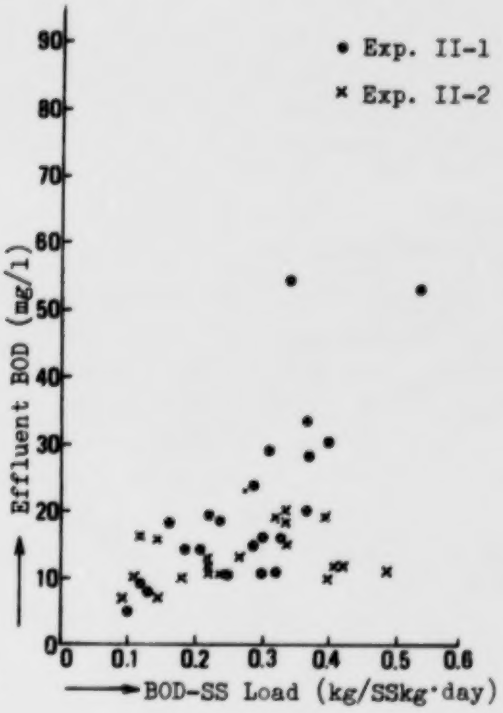


Fig. II-4 Relation between BOD-SS Load and BOD in the Effluent



As shown by Figures II-5 and II-6, the combined treatment of the ordinary municipal sewage and Route B sewage brings about a very stable, efficient result. In case that the aeration period is 2.5 hours and that the BOD-SS load is in a wide range of 0.1 to 0.5 kg/SSkg·day, the BOD removal is 95% up and BOD in the effluent is less than 20 mg/l.

In other words, if only the high load sewage of Route B is treated by the pure oxygen aeration process so as to purify the effluent to the degree of the discharge water quality standard at the peak of the worst inflow, it requires approximately 3 hour aeration. (in this case, aeration period on the average flow is 3.8 hours.) If, however, the ordinary municipal sewage is mixed with Route B sewage in the even ratio, the aeration period on the average flow can be shortened to 2.5 hours or less. Further it is found by the experiment that the BOD-SS load should range from 0.2 to 0.4 kg/SSkg·day for the most stable, efficient treatment.

When the temperature of the sewage is high, SVI goes down to 40 or so and the return sludge suspended solids reaches 30,000 to 38,000 mg/l and further MLSS can be maintained as high as 6,000 to 7,000 mg/l.

Fig. II-5 Relation between BOD-SS Load and BOD Removal

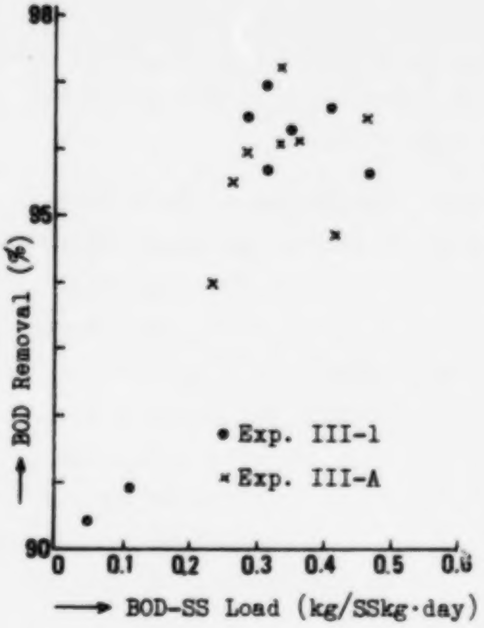
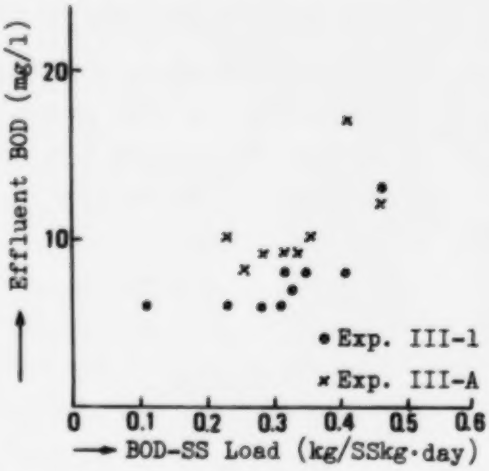


Fig. II-6 Relation between BOD-SS Load and BOD in the Effluent



(2) S S

As shown by Figure II-7, in case that only Route B sewage is treated, the SS removal largely fluctuates between 20 and 90% and is quite unstable disregarding the BOD-SS load. As shown by Figure II-8, however, in case that the municipal and Route B sewage are mixed, the result is stable with the removal being 75% and over and SS in the effluent being less than 20 mg/l.

Fig. II-7 Relation between BOD-SS Load and SS Removal

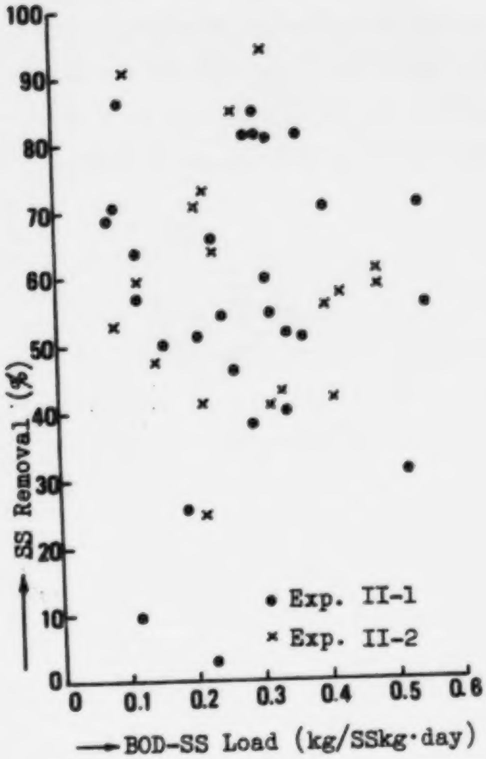
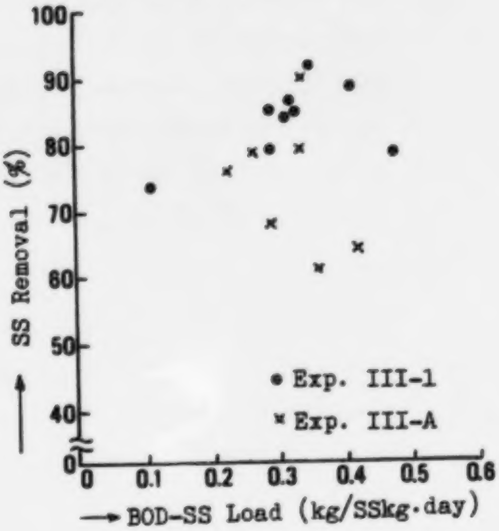


Fig. II-8 Relation between BOD-SS Load and SS Removal



(3) Consumption of Oxygen

To remove 1 kg of BOD, 1.0 to 2.0 kg of oxygen is required. Compared with the treatment of ordinary municipal sewage, this amount of oxygen is fairly big. As for the relation between DO in mixed liquor and the treatment efficiency, if the DO at the 1st section of the aeration tank is low and that of the 4th section is high, the treatment is done efficiently, but if the DO of the 1st section is as stipulated and that of the 4th section lowers, the treatment deteriorates.

(4) Characteristics of Activated Sludge

The flock of activated sludge is microscopic and the zoogloea that forms the flock does not branch out but is spherical. When the sludge shows a good precipitation, the SVI is around 40 and the sludge precipitation rate is as big as 3.0 to 4.0 cm/min, and the return sludge suspended solids is as high as 30,000 to 38,000 mg/l.

(5) Production of Excess Sludge

The production of excess sludge solids is 0.38 kg per 1 kg of removed BOD and 68% of it is volatile solids. The amount of sludge solids is 87g per 1 cubic meter of the sewage treated on the average. Usually the volume of sludge with moisture is 0.34% of the sewage treated. At the ordinary air system it is said to be 1% but by this process it is only 1/3.

III. Conclusion

Through various experiments, which were conducted to treat Route B sewage to the degree of discharge water quality standard, the following differences were made clear between the pure oxygen aeration process and the ordinary air system.

1) The ordinary air system requires 10 hours of aeration period and over (by the two stage treatment process it is 7 hours), while the pure exygen process requires only 4 hours.

2) The BOD-SS load is 0.15 kg/SSkg·day for the ordinary air system, the same is 0.2 to 0.4 kg/SSkg·day for the pure oxygen process.

3) The air supplied is approximately 15 m^3 air/m^3 sewage at the ordinary air system, while at the pure oxygen aeration process the amount of oxygen consumed for the removal of 1 kg of BOD is 1.0 to 2.0 kg. Consequently, compared with the ordinary air system, the pure oxygen process.

   (a) well removes BOD at the higher BOD-SS load and

   (b) produces less excess sludge.

Therefore, it is now evident that the pure oxygen aeration process in quite useful for the treatment of the highly polluted sewage with a largely variable load, such as the one inflowing to Kisshoin Sewage Treatment Plant of Kyoto.

Third US/JAPAN Conference

STUDIES ON ADVANCED WASTE TREATMENT

presented by
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STUDIES ON ADVANCED WASTE TREATMENT

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1. Laboratory tests

1.1 Lime precipitation in municipal wastewater

At the Second U.S.-Japan Conference on Sewage Treatment Technology held in Washington, D.C. and Cincinnati, Ohio in 1972, some of the results of laboratory test concerning with lime precipitation in municipal waste water along with an outline of a pilot plant installed at the Shita-machi Sewage Treatment Plant in Yokosuka and some of pilot-scale investigation results were reported.

These results are briefed as follows:

a. In pH of wastewater was raised up more than 10.5 by lime dose, the concentration of total phosphorus in the supernatant could be reduced to lower than 0.5 mg/l regardless the wastewater, that is, raw sewage, primary or secondary effluent.

b. The lime dose required to raise pH of the wastewater more than 11.0 was as shown in Table 1.1.

c. The effects of the reduction of phosphorus by lime precipitation on the concentration of magnesium in wastewater were as follows;

   While the reduction of metaphosphate depended on the reduction of magnesium hydroxide, the reduction of orthophosphate was not related to the reduction of magnesium hydroxide.

d. Using an X-ray diffraction method, it was disclosed that the final products of calcium and phosphate reaction in lime precipitation process are calcium hydroxylapatite $[Ca_5(OH)(PO_4)_3]$.

e. Scales of calcium carbonete sampled from inside of some tanks at the Yokosuka pilot plant was identified by both electronic microscopy and X-ray diffractmetry to be a mixture of aragonite and calcite. These scales assumed a slab form on the inside wall and a granular shape on the water in the tanks.

The Ministry of Construction gave birth to a project team comprising engineers of eight municipalities which are considered to implement the advanced waste treatment of their own municipal wastewater in the future. It is called "Joint Working Group on Advanced Wastewater Treatment Technology" and undertakes the collection and exchange of information, laboratory and pilot plant tests data gathering and discussion among members over findings. From Tokyo Metropolitan Government, Yokohama City, Nagoya City, Kyoto City, Nagoya City, Kyoto City, Osaka City, Kobe City, Kita-kyushu City and Yokosuka City have come two engineers each to participate in JWGAWTT. From April, 1972 to March, 1973, participate in JWGAWTT. From April, 1972 to March, 1973, the project team members had been mostly involved in laboratory tests of lime precipitation using primary influent and secondary effluent of their own facilities. During this one-year period, the laboraty tests were conducted twice - one in Summer and one in Winter. In the laboratory tests, the conditions were normalized that the flush mixing was set at 150 rpm for 5 min., flocculation at 30 rpm for 25 min. and the settling for 30 min.

The results of analyses of the samples subjected to the laboratory tests are given in Tables 1.2 and 1.3.

As will be clear from tables, the concentration of total phoshorus in the primary influent was in the range of 1.34 mg/l to 69.8 mg/l, while 80 per cent of the samples of influent showed a concentration of total phosphorus of less than 10 mg/l. On the other hand, the concentration of total phosphorus in the secondary effluents ranged from 0.37 mg/l to 5.38 mg/l, whereas 72 per cent of the samples of the secondary effluents showed a concentration of total phosphorus of not more than 1 mg/l. These values are by far lesser than those reported in the United States.

The test results are summarized below:

a. All of the samples consisting of primary influents and secondary effluents were normalized to have a pH value of 10.9 to 11.5

before procipitation. In this case, 43 per cent of the samples less than 0.4 mg/l of total phosphorus concentfation in the supernatant. On the other hand, the concentration of total dissolved phosphorus in the supernatant obtained from 68 per cent of the samples was less than 0.4 mg/l.

In the tests, there were found no definite relationship between the concentration of total phosphorus in the samples and the concentration of total phosphorus and total dissolved phosphorus in the supernatants.

b. The concentration of total phosphorus and the concentration of total dissolved phosphorus, both in the supernatant were compared to each other by the same treatment facilities.

It was disclosed that the lime-precipitated secondary effluent generally showed lower concentration in both total phosphorus and total dissolved phosphorus than the limeprecipitated primary influent. (As regards the total phosphorus, 13 out of 18 samples showed lower values, and as regards total dissolved phosphorus, 14 out of 15 samples showed lower values.)

Thus, it was concluded that in order to reduce the concentration of total phosphorus the lime precipitation of the secondary effluent should be preceded by biological treatment of the primary influent. (See Table 1.4)

c. The concentration of total phosphorus and the concentration of the total dissolved phosphorus, both in the supernatant, were compared to each other by treatment plant and by season. The comparative study revealed that the tests conducted in the summer when water temperature is high goes a long way toward reducing the concentration of both the total phosphorus and total dissolved phosphorus rather than the tests in the winter when water temperature is low.

(As regards the total phosphorus, 13 out of 18 samples showed lower values, and as regards the total dissolved phosphorus 15 out of 18 samples showed lower values.) (See Table 1.5)

d. It was also made clear that the lime dose was necessary to raise pH of the primary influent and secondary effluent up to 11.0 changes with the buffer capacity of wastewater samples. M-alkalinity of the samples was 50 to 200 mg/l, and the lime dose required was 100 to 400 mg/l.

e. Primary influent and secondary effluent showed a magnesium concentration of not more than 60 mg/l. The magnesium in wastewater could be transformed into magnesium hydroxide by adjusting the samples' pH up to 11.0.

   Magnesium hydroxide affected as a flocculation aid, achieving a substantive result in removing total suspended phosphorus in lime-precipitated supernatant.

   It was found that the residue of the total suspended phosphorus in lime-precipitated supernatant is related to the concentration of magnesium hydoroxide produced.

f. Lime-precipitated sludge being rich with magnesium hydroxide presents a low settleability, and increases its volume. (See Fig. 1.1)

g. The increase in pH is depressed by the increase in magnesium concentration in wastewater.

   In order to carry out lime precipitation of wastewater containing much magnesium, the lime dose should be enough to cover up OH^- which is required for the conversion of magnesium-ion into magnesium hydroxide.

h. Increase in calcium concentration in wastewater due to lime dosage varied largely in the range of 20 to 140 mg/l. (lime dose: 400 mg/l)

   This increase had a correlation with M-alkalinity in wastewater; the higher the M-alkalinity was, the lower the increase in calcium in the supernatant resulted.

i. The reduction of organic matter in wastewater by lime precipitation was 30 to 70% for the primary influent and 20 to 50% for the secondary effluent, both in $KMnO_4$-COD(COD_{Mn}) index.

1.2 Investigations now in progress

a. Chemical precipitation of municipal wastewater by metal salts. JWGAWTT and the Public Works Research Institute are carrying cut some laboratory tests of chemical precipitation of removing phosphorus, suspended solids and organic matter by making dose of three kinds of coagulant - aluminum sulfate, ferric chloride and a mixture of aluminum salt and ferric salt (which is available on market and costs less than aluminum sulfate).

b. Reduction of ammoniacal nitrogen by break-point chlorination. In the search for the design criteria of a break-point chlorination process to be built in the Kyoto advanced waste treatment pilot plant, the experimental work has been pushed forward.

1.3 Aspects of future studies

a. With a laboratory lime recalcining furnace installed at the Public Works Research Institute, experimental investigations will be conducted as to recovery and reuse of dewatered lime-precipitated sludge. This furnace will be a modification of a multiple hearth furnace of which only a single hearth is taken out.

b. Also, the same furnace will be used for the regeneration of exhausted granular activated carbon laboratory test.

c. Fundamental investigations on reverse osmosis are planned to be carried out by making use of a flat plate type laboratory use reverse osmotic equipment.

d. For the reduction of phosphate and nitrate in wastewater, ion-exchange method will be examined.

Table 1.1 Lime Dosage to raise pH up to 11.0

mg l^{-1} as $Ca(OH)_2$

Item	Raw Sewage with Digestor Supernatant	Raw Sewage	Primary Sedimentation Tank Effluent	Secondary Sedimentation Tank Effluent
Range	172 ~ 396	135 ~ 450	146 ~ 401	146 ~ 296
Average	293	323	255	235

Table 1.2 Primary Influent Quality Subject to the Laboratory Tests

Season	City	Plant	Water Temperature (°C)	pH	P (mg/1)	P.D (mg/1)	COD$_{Mn}$ (mg/1)	Turbidity[1)] (mg/1)	Ca (mg/1)	Mg (mg/1)	M-Alkalinity (mg/1)
Summer	Tokyo	Ochiai	18.9	7.50	4.46	4.10	85.0	111	27.2	8.5	135
	Yokohama	Chubu	21.5	7.39	4.18	2.92	43.3	116	38.5	43.0	141
		Hokubu	21.5	7.20	2.18	0.160	37.7	71.1	44.0	12.0	154
	Yokosuka	Uwa-Machi	22.0	7.40	1.84	1.02	45.1	22.0	34.0	23.1	111
		Shita-Machi	22.6	7.25	3.98	2.60	77.0	62.0	140	382	98.5
	Nagoya	Meijo	22.0	7.14	8.59	6.85	77.5	396	20.0	42.7	113
		Chitose	21.0	7.00	4.33	1.57	34.2	50.0	70.0	96.0	132
	Kyoto	Toba	23.5	7.31	3.23	2.12	74.1	-	29.4	9.2	81.5
		Kisshoin	23.3	7.37	2.61	2.06	53.4	-	24.7	7.4	124
	Osaka	Nakahama-Nishi	23.0	6.90	10.0	2.66	106	272	49.0	4.9	87.5
		Sumiyoshi	23.8	7.28	26.9	2.13	170	975	48.5	15.2	144
	Kobe	Chubu	23.2	8.34	7.90	5.60	109	260	41.0	67.0	20.1
		Higashinada	23.5	7.30	5.30	3.40	62.6	90.0	36.3	46.4	174
	Kita-Kyusyu	Hiakari	22.7	7.40	23.3	3.26	445	1,300	54.4	43.7	243
		Kogozaki	22.0	6.95	4.02	2.61	46.6	85.0	39.5	13.8	155
	Mean		22.3	7.32	7.52	2.87	97.8	293	46.4	54.3	128
	Range		18.9-23.8	6.90-8.34	1.84-26.9	0.160-6.85	34.2-445	50.0-1,300	20.0-140	4.9-382	20.1-243
Winter	Tokyo	Ochiai	10.0	7.45	5.49	4.56	143	170	14.9	12.7	90.5
	Yokohama	Chubu	15.6	7.70	4.68	3.88	57.0	268	31.0	30.7	148
	Yokosuka	Shita-Machi	15.0	7.67	1.93	1.77	22.0	50.5	-	-	57
	Nagoya	Meijo	12.5	8.40	10.0	9.76	144	540	9.6	6.1	220
	Kyoto	Toba	14.3	7.52	1.34	1.12	53.3	-	-	-	168
	Osaka	Nakahama-Nishi	15.3	7.28	9.80	3.30	115	343	14.8	5.8	86.0
		Sumiyoshi	16.0	7.30	69.8	3.18	322	1,824	39.9	30.1	197
	Kobe	Chubu	13.8	7.60	9.80	7.55	212	200	24.5	21.7	181
	Kita-Kyusyu	Hiakari	14.0	8.55	6.84	1.15	148	750	41.0	41.9	198
		Kogozaki	13.7	7.16	4.30	0.900	248	1,800	22.0	9.30	151
	Mean		14.1	7.66	12.4	3.72	146	661	24.7	19.8	150
	Range		10.8-16.8	7.16-8.40	1.34-69.8	0.900-9.76	22.0-322	50.5-1,824	9.6-41.0	5.8-41.9	57-220
Mean			-	7.46	9.47	3.21	117	444	38.9	42.3	137
Range			-	6.90-8.40	1.34-69.8	0.160 9.76	22.0-445	50.0-1,824	9.6-140	4.9-382	20.1-243

1) Turbidity in mg·kaoline/l

* Samples collected at 9:30 a.m.

Table 1.3 Secondary Effluent Quality Subject to the Laboratory Tests

Season	City	Plant	Water Temperature (°C)	pH	P (mg/1)	P.D (mg/1)	COD_{Mn} (mg/1)	Turbidity[1] (mg/1)	Ca (mg/1)	Mg (mg/1)	M-Alkalinity (mg/1)
Summer	Tokyo	Ochiai	20.3	7.00	0.780	0.670	16.5	-	25.6	5.8	99.0
	Yokohama	Chubu	22.2	7.05	1.04	0.955	5.5	4.4	40.0	34.0	72.3
		Hokubu	22.0	7.30	1.43	1.34	6.8	3.1	31.3	10.0	65.0
	Yokosuka	Uwa-Machi	22.5	7.40	0.450	0.389	3.5	1.9	30.0	12.2	74.0
		Shita-Machi	22.0	7.35	0.756	0.679	7.4	1.4	156	484	97.0
	Nagoya	Meijo	21.5	7.21	1.47	0.718	25.3	46.4	12.8	17.0	83.0
		Chitose	21.0	7.10	0.260	0.240	8.2	3.0	76.0	161	100
	Kyoto	Toba	23.6	7.55	0.480	0.480	14.0	-	26.9	8.4	77.3
		Kisshoin	23.3	7,82	0.548	0.350	36.8	-	23.3	8.1	89.9
	Osaka	Nakahama-Nishi	23.0	6.80	1.33	0.888	16.8	21.5	22.0	3.4	75.0
		Sumiyoshi	23.7	7.31	0.950	0.320	17.3	30.6	48.5	17.4	169
	Kobe	Chubu	24.9	6.88	5.38	4.28	32.8	40.0	52.6	115	77.3
		Higashinada	23.8	7.80	3.38	2.60	17.6	14.0	34.3	28.4	186
	Kita-Kyusyu	Hiakari	22.9	7.03	2.90	2.74	5.3	15.0	60.0	31.6	83.0
		Kogosaki	21.3	7.00	1.62	1.42	5.0	7.5	30.2	7.2	38.0
	Mean		22.5	7.24	1.52	1.20	14.6	15.7	44.6	62.9	92.4
	Range		20.3-24.9	6.80-7.82	0.260-5.38	0.240-4.28	3.5-36.8	1.9-46.4	12.8-156	3.4-484	38.0-186
Winter	Tokyo	Ochiai	16.3	7.31	0.370	0.281	11.0	11.0	15.3	10.6	83.5
	Yokohama	Chubu	14.5	7.50	2.60	2.40	9.9	11.0	35.3	37.3	97.5
	Yokosuka	Shita-Machi	15.8	7.08	0.713	0.690	4.8	2.5	-	-	39.5
	Nagoya	Meijo	13.6	7.00	4.16	3.60	23.6	50.8	-	-	147
	Kyoto	Toba	14.7	7.05	0.969	0.908	22.8	-	-	-	166
	Osaka	Nakahama-Nishi	14.5	7.38	3.23	2.28	20.4	26.3	16.7	2.9	81.0
		Sumiyoshi	13.8	7.52	0.322	0.119	14.3	17.8	40.8	30.4	123
	Kobe	Chubu	17.1	7.00	3.06	2.86	21.9	20.0	25.1	32.8	168
	Kita-Kyusyu	Hiakari	12.6	7.30	0.859	0.782	11.3	15.2	68.1	33.3	213
		Kogosaki	14.6	7.13	1.62	1.40	11.9	21.2	20.4	9.9	105
	Mean		14.3	7.27	1.79	1.53	15.2	19.5	31.7	22.5	122
	Range		11.3-17.1	7.00-7.52	0.370 4.16	0.119-3.60	4.8-23.6	2.5-50.0	15.3-68.1	2.9-37.3	39.5-213
Mean			-	7.25	1.63	1.33	14.8	17.3	40.1	50.0	104
Range			-	6.80-7.82	0.260-5.38	0.119-4.28	3.5-36.8	1.9-50.8	12.8-156	2.9-484	38.0-213

1) Turbidity in mg·kaoline/1

* Samples collected at 9 = 30 a.m

Table 1.4 Phosphorus and Dissolved Phosphorus in the Lime Coagulated Supernatants of Primary Influent and Secondary Effluent (Summer Samples)

Plant	pH	$Ca(OH)_2$ (mg/l)	P (mg/l)	P.D (mg/l)	Secondary Effluent/ Primary Influent	
					as P	as P.D
Ochiai	11.70	400	0.25	0.13	0.28	0.15
	11.70	400	0.07	0.02		
Chubu (Yokohama)	11.00	400	0.052	0.020	0.58	0.40
	10.80	200	0.030	0.008		
Hokubu	11.41	400	0.062	0.028	0.66	0.36
	11.30	300	0.041	0.010		
Uwa-Machi	11.51	400	0.100	0.052	0.59	0.46
	11.42	400	0.059	0.024		
Shita-Machi	11.21	400	0.164	0.137	0.20	0.12
	10.65	400	0.033	0.017		
Meijo	11.71	400	0.870	0.160	0.32	1.39
	11.66	400	0.278	0.222		
Chitose	10.9	400	0.20	0.11	0.45	0.82
	10.5	400	0.090	0.09		
Toba	12.22	400	0.350	0.165	0.57	0.92
	12.25	400	0.201	0.152		
Kisshoin	12.16	400	0.770	0.205	0.25	0.44
	12.02	400	0.195	0.090		
Nakahama-Nishi	11.3	400	0.72	0.50	0.35	0.28
	11.2	400	0.25	0.14		
Sumiyoshi	11.41	400	0.66	0.50	0.03	0.70
	11.04	300	0.023	0.07		
Chubu (Kobe)	11.44	400	0.148	0.141	0.45	0.29
	11.14	300	0.067	0.041		
Higashinada	11.62	400	0.095	0.066	1.39	0.45
	11.42	400	0.132	0.030		
Hiakari	10.52	400	0.822	0.254	0.55	0.41
	10.24	100	0.456	0.104		
Kogozaki	11.45	400	0.208	0.164	1.19	0.24
	11.40	200	0.248	0.040		

* Upper figures in colomns : primary influent

* Lower figures in colomns : secondary effluent

Table 1.5 Phosphorus and Dissolved Phosphorus in the Lime Coagulated Supernatants of Summer Samples and Winter Samples

Plant	Primary Influent						Secondary Effluent					
	pH	$Ca(OH)_2$ (mg/l)	P (mg/l)	P.D (mg/l)	Summer/Winter		pH	$Ca(OH)_2$ (mg/l)	P (mg/l)	P.D (mg/l)	Summer/Winter	
					as P	as P.D					as P	as P.D
Ochiai	11.70 11.88	400 300	0.25 0.310	0.13 0.212	0.80	0.61	11.70 11.84	400 300	0.07 0.093	0.02 0.047	0.75	0.43
Chubu (Yokohama)	11.00 11.30	400 300	0.052 0.28	0.020 0.23	0.18	0.09	11.40 11.8	400 400	0.008 0.03	0.0 0.03	0.27	0
Shita-Machi	11.21 11.24	400 200	0.164 0.408	0.137 0.325	0.40	0.42	10.65 11.65	400 400	0.033 0.068	0.017 0.040	0.49	0.43
Meijo	11.34 11.52	300 400	4.21 4.16	0.180 3.12	1.01	0.06	11.48 11.60	300 400	0.340 1.680	0.216 1.480	0.20	0.15
Toba	11.03 11.42	200 300	1.45 0.806	0.485 0.570	1.79	0.84	11.10 11.43	150 400	0.284 0.580	0.280 0.452	0.49	0.62
Sumiyoshi	11.41 11.44	400 400	0.66 0.945	0.50 0.688	0.69	0.73	11.45 11.55	400 300	0.05 0.053	0.03 0.049	0.94	0.61
Nakahama-Nishi	11.30 11.57	400 200	0.72 1.81	0.50 0.635	0.39	0.79	11.2 11.50	400 150	0.25 2.120	0.14 0.900	0.12	0.16
Hiakari	10.52 11.15	400 300	0.822 0.323	0.254 0.056	2.54	4.54	10.68 10.79	150 400	0.205 0.091	0.020 0.062	2.25	0.32
Kogozaki	11.45 11.73	400 300	0.208 0.409	0.164 0.158	0.50	1.04	11.14 11.30	150 400	0.840 0.045	0.166 0.050	18.69	3.32

* Upper figures in columns : Summer samples (mean water temp. 22.4°C)

* lower figures in columns : winter samples (mean water temp. 14.2°C)

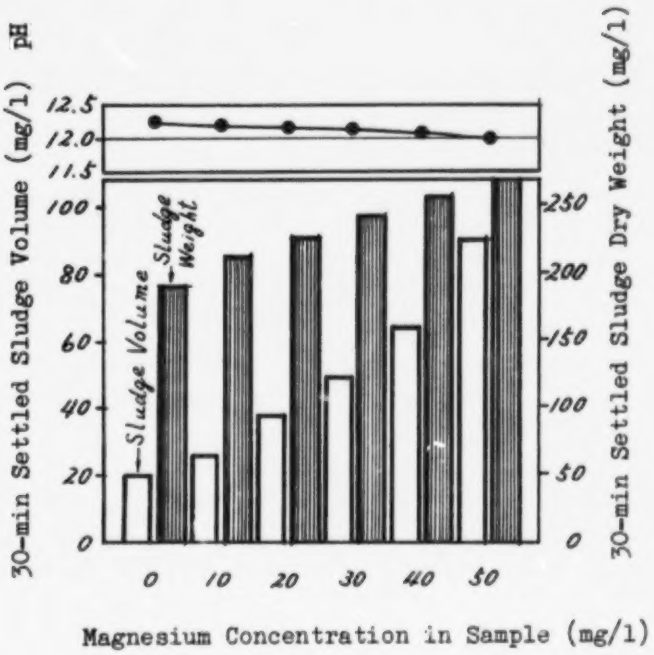


Fig. 1.1 Influence of Magnesium on Sludge Volume and pH of Lime Treated Wastewater ($Ca(OH)_2$ = 400 mg/1)

2. Yokosuka pilot plant studies

2.1 Operating conditions of pilot plant

A flow diagram of the pilot plant is shown in Fig. 2.1, and its operating conditions are given in Table 2.1.

The lime dose was 300 mg/l as hydrated lime during the period from June 22, 1972 to March 22, 1973 (hereinafter referred to as "A" series) and 1,000 mg/l as hydrated lime during the period from May 9, 1973 to October 13, 1973 (hereinafter referred to as "B" series).

pH of the wastewater after adjustment by lime dose was 10.45 on the average in "A" series and 10.95 on the average in "B" series.

2.2 Characteristics of pilot plant influent

Shita-machi Sewage Treatment Plant, Yokosuka, where the pilot plant under discussion is installed, takes in domestic sewage as well as wastes from marine products processing factories which use much seawater. Also, seawater permeates into the sewar pipes laid along the coastal line. Therefore, the influent to the treatment plant contains significant amount of seawater, and magnesium concentration is relatively high. In the lime precipitation, this high concentration of magnesium restrains the rise of pH; even when as much as 1,000 mg/l of hydrated lime was dosed, pH of wastewater after pH adjustment was 10.95 on the average.

Yokosuka city is planning to reuse the treated effluent as industrial water in the future, and is in the process of investigating the sewer system reconstruction in order to prevent infiltration of seawater.

The lime precipitation of wastewater containing much magnesium brings about diversified effects due to a large amount of precipitated magnesium hydroxide as already described in Section 1.1. "Lime precipitation in municipal wastewater."

Namely, magnesium hydroxide acts as a coagulant aid, while it retards the rise of pH of wastewater (see Fig. 2.2), forms light flocs with poor settleability, sludge thickening rate and dewaterability, and may cause the accumulation of magnesium in recalcined lime when lime is recovered.

The concentration of magnesium in the wastewater handled by Shitamachi Sewage Treatment Plant is 300 to 400 mg/l as against about 10 mg/l in the typical municipal wastewater in Japan.

2.3 Outline of the test results

Table 2.2 shows the mean values and ranges of influent and effluent quality and average removal concerning "A" and "B" series at the pilot plant.

Average suspended solids concentration of the influent were 5.4 mg/l in "A" series and 2.6 mg/l in "B" series. The turbidity was 5.8 mg/l and 3.6 mg/l, respectively.

On the other hand, suspended solids and turbidity of the lime sedimentation tank effluent were much more than those in the influent; namely, the suspended solids were 12.3 times as much in "A" series and 13.9 times as much in "B" series. This is because the precipitates, such as calcium carbonate and magnesium hydroxide, could not be settled down to the level of the suspended solids of the influent.

In the calcium carbonate settling tank, a considerable portion of suspended solids and turbidity was removed. In "A" series, suspended solids and turbidity in the effluent of the calcium carbonate settling tank was 5.0 per cent and 7.8 per cent of those in the effluent of the lime sedimentation tank.

In the dual-media filter, both suspended solids and turbidity were removed quite well. It is worthwhile that the removal of suspended solids was almost 100 per cent for both "A" and "B" series. The turbidity of the effluent of the dual-media filter was 0.3 mg-Kaoline/l for "A" series and 0.08 mg-Kaoline/l for "B" series.

The total phosphorus contents of the influent was 1.21 mg/l for "A" series and 1.43 mg/l for "B" series. The concentration of total phosphorus of the lime sedimentation tank effluent was 0.344 mg/l for "A" series and 0.125 mg/l for "B" series. pH of the lime sedimentation tank effluent was 10.28 and 10.81 for "A" and "B" series, respectively.

The removal of total phosphorus by lime precipitation in "A" series was lower than that in "B" series. The total phosphorus in the effluent of the calcium carbonate settling tank was 0.114 mg/l or 90.5 per cent removal in "A" series. And the total phosphorus in the effluent of the dual-media filter was 0.102 mg/l and 0.04 mg/l or 91.6 per cent and and 97.1 per cent for "A" and "B" series, respectively.

The ratio of orthophosphate to the total phosphorus was 87 per cent for both "A" and "B" series so long as the pilot plant influent was concerned. On the contrary, the ratio of orthophosphate in the effluent of dual-media filter was 64 per cent and 77 per cent for "A" and "B" series, respectively.

COD_{Cr} and TOC in the pilot plant influent was 40.2 mg/l and 40.4 mg/l for "A" series and 41.9 mg/l and 49.2 mg/l for "B" series, respectively. COD_{Cr} and TOC were removed little by little in each unit process. The over-all removal of COD_{Cr} throughout the whole processes was 44.3 per cent and 50.0 per cent for "A" and "B" series, respectively, and that of TOC was 25.3 per cent and 28.9 per cent, respectively, with the result that the removal of COD_{Cr} was larger than that of TOC.

The regression line of COD_{Cr} with respect to TOC was found as follows,

$$COD_{Cr} = 0.673 \cdot TOC + 5.90$$

The correlation coefficient was 0.761.

The total nitrogen in the pilot plant influent was 7.12 mg/l and 7.83 mg/l for "A" and "B" series, respectively.

The over-all removal of the total nitrogen throughout the whole processes was 10.8 per cent and 30.3 per cent for "A" and "B" series, respectively.

The ratio of ammoniacal nitrogen (NH_3-N) to the total nitrogen in the pilot plant influent was 19 per cent and 50 per cent for "A" and "B" series, respectively. The ratio was 7.6 per cent and 36 per cent for "A" and "B" series, respectively, in the dual-media filter effluent.

The removal of ammoniacal nitrogen by the ammonia stripping tower was as low as 31.6 and 39.9 per cent for "A" and "B" series, respectively. This may be ascribable to the fact that the air-liquid ratio was as small as 700 to 2,000. The air-liquid ratio was first designed to be in the range of 1,000 to 4,000 $M^3 \cdot$air/$m^3 \cdot$liquid, but the actual head loss in the tower exceeded the estimated value to reduce it to the aforesaid value. At the air-liquid ratios, 700 and 1,500, the removal of ammoniacal nitrogen was 20 per cent and 50 per cent, respectively. The variation of liquid temperature was in the range of 15°C to 23°C, during the period of the experiment, and does not seem to have significant effect on the removal of the ammoniacal nitrogen.

2.4 Considerations to phosphorus removal

The mean values of total phosphorus and their standard deviations are related to the mean liquid temperatures of the dual-media filter effluent for "A" series as shown in Fig. 2.3.

The figure indicates that with the lime dose of 300 mg/l, the higher the liquid temperature is, the lower becomes the concentration of the total phosphorus in the dual-media filter effluent and also the smaller becomes the fluctuations.

Comparing "A" series (mean liquid temperature: 22.7°C) with "B" series (mean liquid temperature: 13.0°C) in the figure, it appears that an increase of about 10 degrees centigrade in liquid temperature acts to reduce the total phosphorus from 0.128 mg/l to 0.050

mg/l and to increase the standard deviation from 0.081 mg/l to 0.25 mg/l.

In "A" series, where lime dose was 300 mg/l, the effects of the overflow rate of lime sedimentation tank and sludge recirculation ratio upon the removal of phosphorus are shown in Table 2.3. As may be seen in Table 2.3, a lowering of the overflow rate of the lime sedimentation tank and the application of sludge recircutation not only reduce the total phosphorus in the effluent of the lime sedimentation tank, but also make its fluctuation smaller.

In "B" series, where lime dose was 1,000 mg/l, there were little or no effects of the overflow rate and sludge recirculation.

As evidenced by the comparison between "A" series and "B" series, the change of lime dose from 300 mg/l to 1,000 mg/l resulted in the reduction of total phosphorus in the effluent both of the lime sedimentation tank and of the dual-media filter. and diminished the fluctuation of the total phosphorus concentration.

2.5 Scale formation of calcium carbonate

In this pilot plant, development of scale of calcium carbonate was noticed. It spreaded from the flush mixing tank to flocculation tank, and was remarkable in winter rather than in summer. Photo 2.1 shows a formation of scale on the flush mixing tank. In winter, development of scale was so serious that the predetermined flow rate for the pilot plant operation could not maintain owing to resultant reduction in the cross-sectional area of a pipe connecting each tank and increase of roughness inside of the pipe. When the scale development was most serious, the predetermined flow rate of 9 m^3/hr could be maintained only for the first 8 days after cleaning, with gradual decrease afterwards, and on the 16th day the flow rate was decreased to about 1 m^3/hr.

Table 2.4 lists the compositions of scale formed in the tanks and pipes of the pilot plant. For the preparation of the table, 10 samples were collected from the extension between the flush mixing tank to the ammonia stripping tower and analyzed.

No significant difference was found in the analytical results of 10 sample.

$CaCO_3$ equivalent of CaO listed in the table was 89.4 per cent, signifying that calcium carbonate accounted for about 90 per cent of scale compounds.

In "A" series (lime dose: 300 mg/l; average pH value of ammonia stripping tower effluent: 10.28; liquid temperature: 10° to 24°C), no scale formation was noticed, except for a slight deposition over the filler surfaces in the ammonia stripping tower. In "B" series, on the other hand, scale formation in "B" series (lime dose: 1,000 mg/l; average pH value of the ammonia stripping tower effluent: 10.81; and liquid temperature: 15° to 27°C) was noticeable; during 5 months of operation, scale deposition over the filler surfaces was 0.5 to 1.5 cm in thickness.

In the lime precipitation-ammonia stripping processes, the most serious problems is considered the scale formation.

Improvements taken so far for the pilot plant against scale problem are as follows.

a. Connection between tanks was changed from pipe system to open channel type for ease of monitoring and removing scale.

b. A part of the open channel was attached with a scale preventive device which applys feeble current and magnetic field. Its functional effects are under study.

2.6 Investigations now in progress

The Investigation now in progress are compared with those in the past in the following.

a. In the past, powdered lime was dosed in the form of slurry, while at present quick lime is slaked and then dosed in the flush mixing tank.

b. In order to improve settleability in the lime precipitation tank, an anionic polymer is dosed as the flocculant aid.

c. Metal salts precipitation process is being studied in parallel with the lime precipitation process.

d. Granular activated carbon adsorption contractors are provided following the dual-media filter.

e. Capacity of ventilator for the ammonia stripping tower is increased to increase air-liquid ratio up to 2,000 to 5,000.

f. Scale preventive device is installed for lime precipitation process.

A part of the results of the experiments now under way is summarized in Table 2.5.

2.7 Aspects of future studies

The facilities and investigations planned for the future are listed below.

a. Rearrangement of piping to transfer metal salts precipitated effluent to one of the existing dual-media filters.

b Installation of control devices which are able to regulate lime and metal salts coagulants dose in response to the changes in the quality fluctuations of influent.

The lime dose will be controlled depending on either pH or alkalinity in the influent, and also metal salts will be controlled depending on either concentration of phosphate in influent, pH of liquid in flush mixing tank or turbidity of sedimentation tank effluent. Control system for the pilot plant is still undecided.

c. Installation of centrifuge for sludge dewatering. Operating conditions of the centrifuge, especially for lime sludge dewatering to separate calcium carbonate from other compounds will be determined.

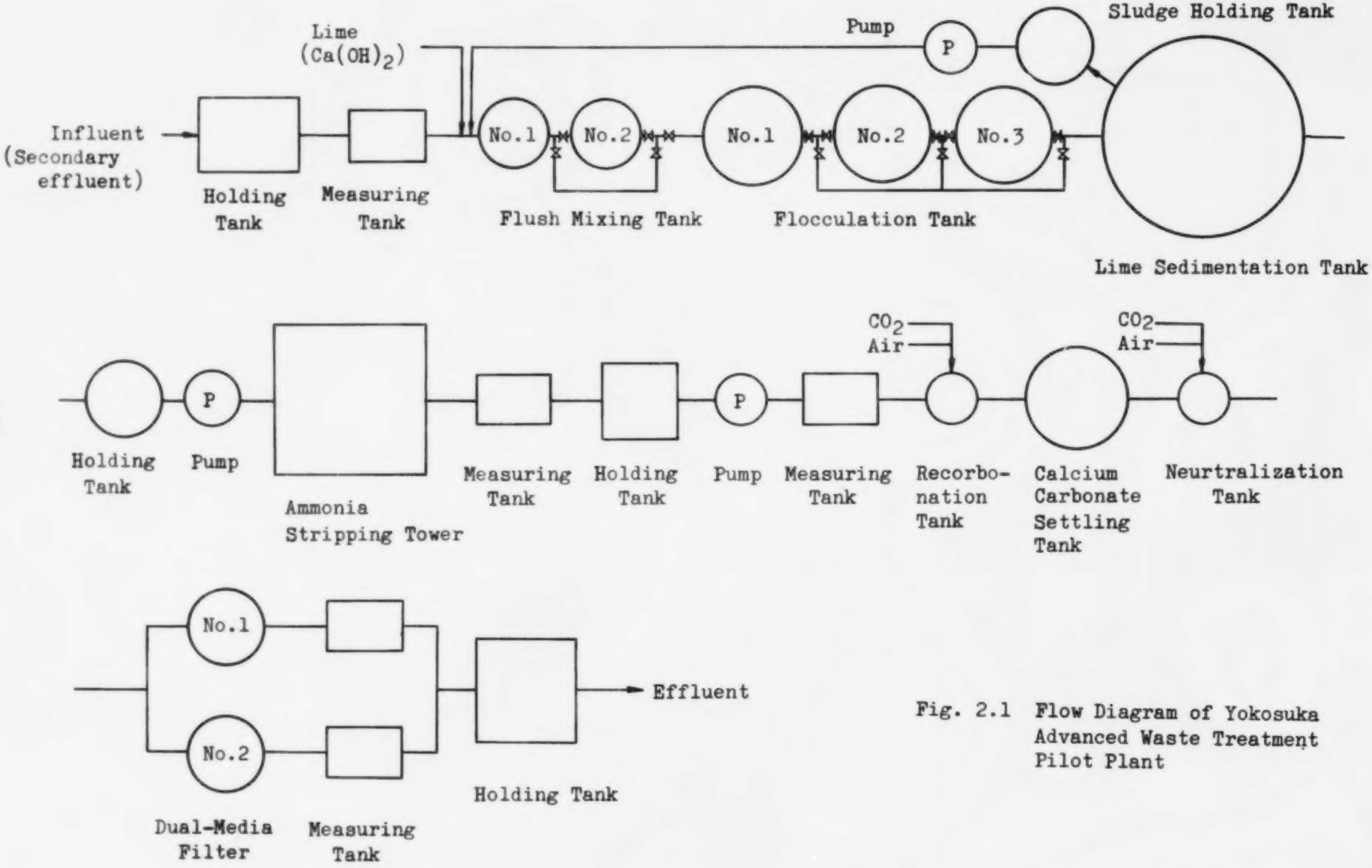


Fig. 2.1 Flow Diagram of Yokosuka Advanced Waste Treatment Pilot Plant

Table 2.1 Operation Conditions of the Pilot Plant

Pilot Plant Influent	Secondary Effluent	
Lime Sedimentation Tank	Flow Rate	6 ~ 9 m^3/h
	Overflow Rate	30 ~ 50 $m^3/m^2/d$
	Detention Time	2.3 ~ 1.5 h
	Sludge Recirculation Ratio	0 ~ 20 %
Ammonia Stripping Tower	Flow Rate	3 ~ 8.5 m^3/h
	Hydraulic Loading	60 ~ 200 $m^3/m^2/d$
	Air-liquid Ratio	700 ~ 2000 m^3.air/ m^3 liquid
Calcium Carbonate Settling Tank	Flow Rate	2.5 ~ 6 m^3/h
	Overflow Rate	30 ~ 80 $m^3/m^2/d$
	Detention Time	1.8 ~ 0.62 h
Dual-Media Filter	Flow Rate	2 ~ 5.5 m^3/h
	Filtration Rate	120 ~ 350 m/d

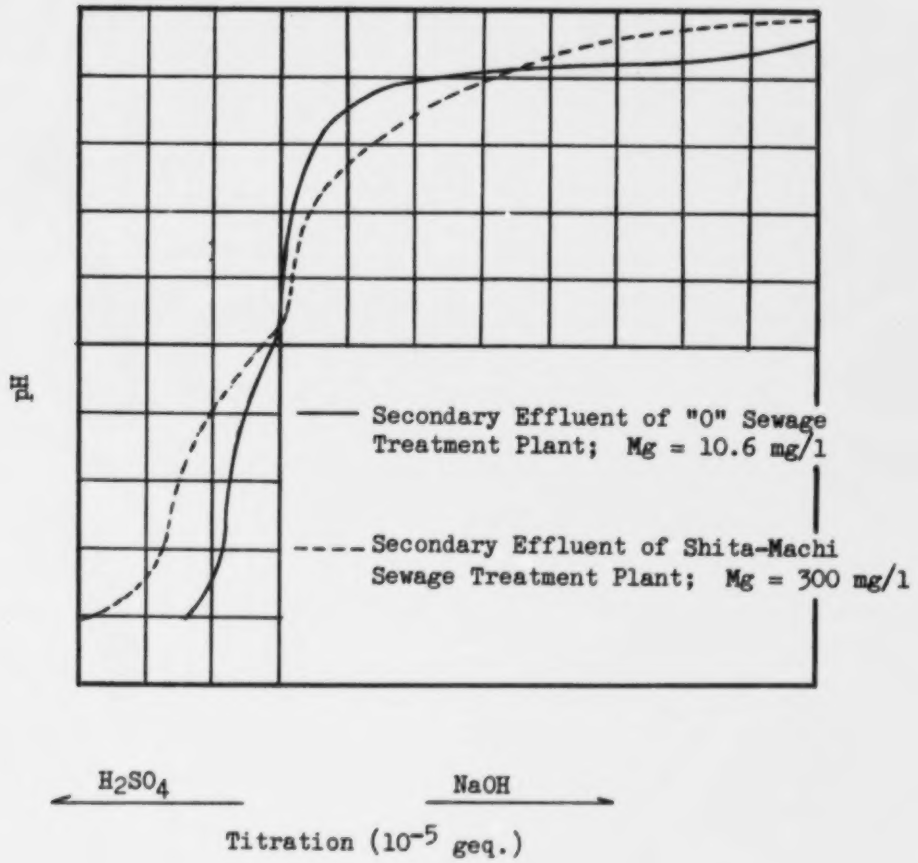


Fig. 2.2 Titration Curves for Wastewaters of Different Magnesium Concentration

Table 2.2 Mean and Range of Water Quality in Each Process, and Accumulated % Removal

		A Series (1972.6.12 - 1973.3.22)					B Series (1973.5.9 - 1973.10.13)				
		Influent (Secondary Effluent)	Lime Sedimentation Tank Effluent	Ammonia Stripping Tower Effluent	Calcium Carbonate Settling Tank Effluent	Dual-Media Filter Effluent	Influent (Secondary Effluent)	Lime Sedimentation Tank Effluent	Ammonia Stripping Tower Effluent	Calcium Carbonate Settling Tank Effluent	Dual-Media Filter Effluent
PH	mean	6.96	10.28	9.90	8.43	7.95	7.65	10.81	10.29	9.65	7.24
	range	6.90~7.90	9.30~11.20	9.20~10.75	6.00~10.60	6.30~10.50	6.90~9.15	10.00~12.40	8.70~12.50	6.38~11.90	5.70~10.90
Turbidity 1) (mg/1)	mean	5.8	39.8	-	3.1	0.3	3.6	23.8	-	6.0	0.08
	range	0.8~33.5	2.0~510	-	0.8~20.5	0 ~21.0	1.5~11.0	3.0~130	-	0.2~45.5	0 ~0.65
	removal2)	-	-	-	46.5	94.8	-	-	-	-	97.8
Suspended Solids (mg/1)	mean	5.4	66.5	-	3.3	0	2.6	36.1	-	6.2	0
	range	0.4~48.0	8.0~400	-	0 ~22.0	0	0.4~8.2	7.0~68.0	-	0 ~40.0	0
	removal2)	-	-	-	38.9	100	-	-	-	-	100
P (mg/1)	mean	1.21	0.344	-	0.114	0.102	1.43	0.125	-	-	0.042
	range	0.403~3.21	0.024~3.28	-	0 ~0.705	0.475	0.702~4.35	0.004~0.696	-	-	0.004~0.201
	removal2)	-	71.5	-	90.5	91.6	-	91.3	-	-	97.1
P,hydro (mg/1)	mean	-	-	-	-	-	1.33	0.102	-	-	0.035
	range	-	-	-	-	-	0.630~2.87	0.006~0.595	-	-	0.006~0.178
	removal2)	-	-	-	-	-	-	92.3	-	-	97.4
P,ortho (mg/1)	mean	1.05	0.352	-	0.075	0.073	1.24	0.086	-	-	0.027
	range	0.503~2.48	0.026~3.68	-	0 ~0.353	0 ~0.394	0.450~2.80	0.004~0.486	-	-	0.002~0.400
	removal2)	-	66.5	-	92.9	93.0	-	93.1	-	-	97.8
TOC (mg/1)	mean	40.4	37.2	-	30.2	30.2	49.2	39.1	-	31.2	35.0
	range	2.4~114	3.3~94.8	-	1.4~86.4	6.2~89.2	6.9~78.9	14.9~96.5	-	12.5~58.1	9.7~70.1
	removal2)	-	7.9	-	25.3	25.3	-	20.5	-	36.6	28.9
COD$_{Cr}$ (mg/1)	mean	40.2	38.1	-	30.9	22.4	41.9	33.6	-	24.6	21.0
	range	4.0~104	4.0~152	-	2.4~99.2	3.2~60.8	13.8~88.9	11.7~68.9	-	9.36 42.0	6.1~54.6
	removal2)	-	2.7	-	23.1	44.3	-	19.8	-	41.3	50.0
T-N (mg/1)	mean	7.12	7.08	-	-	63.9	7.83	-	-	-	5.46
	range	2.12~11.8	2.95~12.3	-	-	6.54~10.5	3.49~9.57	-	-	-	2.13~8.15
	removal2)	-	0.6	-	-	10.3	-	-	-	-	303
NH$_3$-N (mg/1)	mean	1.36	1.33	0.91	-	0.48	3.89	4.14	2.49	-	1.99
	range	0 ~7.41	0 ~6.22	0 ~4.35	-	0 ~3.73	0.51~7.61	1.10~6.03	0.66~3.82	-	0.70~3.38
	removal2)	-	2.2	33.1	-	64.7	-	-	35.9	-	48.8

1) Unit of Turbidity is mg·kaoline/1.
2) % removal is based on influent concentration.
- M-alkalinity of influent were 92 mg/1 (67~119) in "A" series and 64 mg/1 (30~88) in "B" series.
- pH of wastewater after pH adjustment were 10.45 (10.01~10.81) in "A" series and 10.95 (10.00~12.60) in "B" series.
- MLSS in flocculation tank were 994 mg/1 (106~4,180) in "A" series and 1,550 mg/1 (572~3,180) in "B" series.

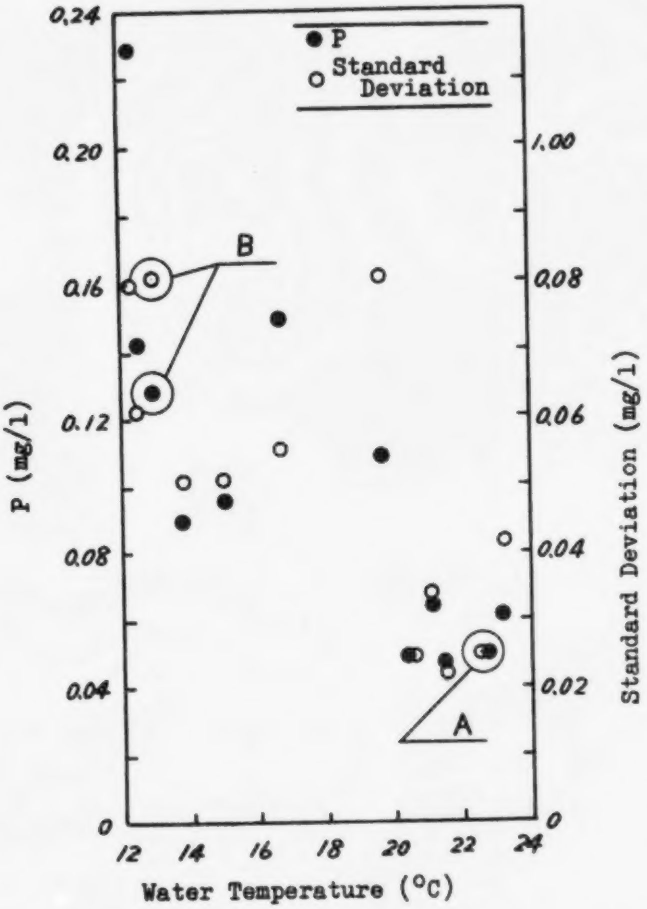


Fig. 2.3 Relationships between Total Phosphorus in Dual-Media Filter and its Standard Deviation, and Water Temperature ("A" series)

Table 2.3 Effects of Overflow Rate of Lime Sedimentation Tank and Sludge Recirculation Rate on Phosphorus Removal

Case	Flow Rate (m^3/h)	Sludge Recirculation Ratio (%)	Overflow Rate ($m^3/m^2.d$)	MLSS in Floccutation Tank (mg/l)		P in Lime Sedimentation Tank Effluent (mg/l)			P in Dual-Media Filter Effluent (mg/l)
				mean	range	mean	Standard deviation	range	
A1	9	0	50	750	300~1,560	0.376	0.146	0.164~0.634	0.050
A2	6	0	33.3	590	140~1,440	0.242	0.107	0.039~0.417	0.048
A3	8	20	44.4	1,600	1,140~2,460	0.210	0.090	0.107~0.394	0.049

* Lime Dose 300 mg/l (as $Ca(OH)_2$)

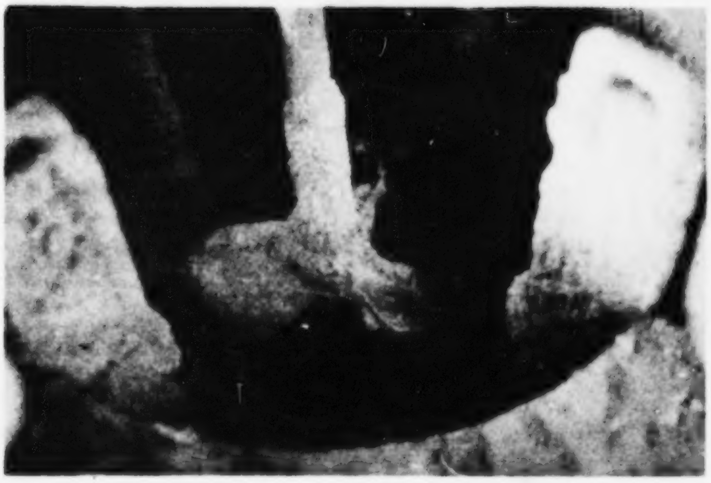


Photo 2.1 Scale on the Flush Mixing Tank

Table 2.4 Composition of Scale

(%)

	Ignition Loss 1)	Impurities 2)	CaO	MgO	Total
Range	45.8~47.2	0.29~1.87	44.3~51.5	2.2~6.4	99.1~102.7
Mean	46.4	0.85	49.5	3.5	100.2

1) at 1,000±5°C; includes adhesive moisture, crystal water and CO_2

2) includes SiO_2 and R_2O_3

3) 49.5% as CaO is equivalent to 89.4% as $CaCO_3$.

Table 2.5 Interin Summary for the Experiments Being Under Way

		Influent (Secondary Effluent)	Lime Sedimentation Tank Effluent	Ammonia Stripping Tower Effluent	Calcium Carbonate Setting Tank Effluent	Dual-Media Filter Effluent	3) Sedimentation Tank Effluent Treated by $Al_2(SO_4)_3$
PH	mean	7.26	10.40	10.06	7.70	7.36	6.85
Turbidity1)	mean4)	5.69	14.5	-	4.70	0.49	5.70
	removal5)	-	-	-	17.3	91.4	-
Suspended Solids	mean4)	5.8	34	-	5.6	0	7.6
	removal5)	-	-	-	3.4	100	-
P	mean4)	1.37	0.322	-	-	0.209	0.433
	removal5)	-	76.5	-	-	84.7	68.4
P,ortho	mean4)	1.17	0.206	-	-	0.174	0.379
	removal5)	-	82.4	-	-	85.1	67.6
TOC	mean4)	28.6	20.9	-	24.5	16.3	22.8
	removal5)	-	26.9	-	14.3	43.0	20.3
$COD_{cr.}$	mean4)	23.4	21.9	-	16.6	12.0	19.9
	removal5)	-	6.4	-	29.1	48.7	15.0
T-N	mean4)	4.86	5.00	-	-	4.28	3.89
	removal5)	-	-	-	-	11.9	20.0

1) Unit of turbidity is mg·kaoline/l

2) lime dose = 300 mg/l as $Ca(OH)_2$

3) mole ratio of Al/P = 3.28

4) mg/l

5) %; based on influent concentration

3. Kyoto pilot plant studies

Advanced waste water treatment plant was partly completed at Toba Seawage Treatment Plant, Kyoto, in February, 1973, and experiments on the down-flow filtration process have been carried out from this March. The down-flow filtration process is designed to remove suspended solids from the conventional activated sludge process effluent.

In Japan, the activated sludge process accounts for 71 per cent of the total number of municipal sewage treatment plants. However, at the final settling tanks of the activated sludge process, organic suspended solids (i.e. biological flocs) may not be removed enough from the effluent.

For this reason, a design manual for the facilities to remove them has been expected to prepare quickly as possible.

To fulfil the demand, Kyoto pilot plant began with the down-flow filtration process for which some information on design and maintenance is available from our own experience of supplying in-plant water.

3.1 Experimental instrumentations and methods of experiments

A pilot filter system was installed, which consists of 3 filters, each having and filter area of 1.2 m^2, a media depth adjustable in the range of 0.6 to 1.0 m, and a maximum filtration head adjustable in the range of 2.4 to 3.0 m. The flow control is carried out by means of a sluice valve, and the filtration head loss is continously measured using an automatic level gauge. Flow rate of effluent is measured continuously by a 30° V-notch weir equipped with an automatic level gauge. The filter is cleaned by fixed surface washing and back washing.

The cleaning wastewater is measured by means of a 90° V-notch weir and sampled when it is necessary.

The filter media size used for the experiments are as listed in Table 3.1. The filter media are packed as illustrated in Table 3.2.

As a first phase experiment, the effects of the ratio of anthracite coal depth to silica sand depth in the dual-media filter on the filtration efficiency and overall economics were examined by comparing the filters No.1 and No.2, along with a comparative study of filters No.2 and No.3 on the effects of the difference between the dual-media system and tri-media system. For the experiments, a declining filtration method was used with the maxium filtration head set at 3.0 m. The initial filtration flow-rate of filters was set at 120 $m^3/m^2 \cdot d$, 240 $m^3/m^2 \cdot d$, 360 $m^3/m^2 \cdot d$, 480 $m^3/m^2 \cdot d$ and 540 $m^3/m^2 \cdot d$, respectively.

The surface washing rate and its duration time for the fixed surface washing were set at 0.15 to 0.2 $m^3/m^2 \cdot$ min. and 6 to 10 min., , respectively, and the expansion rate of filter media and washing time for the back washing at 20 to 25 per cent and 6 to 12 min., respectively. In case the expansion rate for the back washing was 25 per cent, the back washing rates of filters No.1, No.2 and No.3, were 0.78 $m^3/m^2 \cdot$ min., 0.67 $m^3/m^2 \cdot$ min. and 0.70 $m^3/m^2 \cdot$ min., respectively. The fixed surface washing and back washing were overlapped in the time range of 1 to 7 min.

3.2 Relationships between run-length and filtration flow-rate

The relationships between the run-length and filtration flow-rate are shown in Fig. 3.1.

The run-length of No.1 filter was such less than that of filters No.2 and No.3.

The run-length of filter No.2 showed a tendency to become a little longer than that of filter No.3. Namely, it is found that the thickness of anthracite coal medium has a great influence on the run-length, and that the thickness of anthracite coal media to total media depth governs the economics of the filter process, accordingly.

In the declining filtration method used for the experiments, the decline in the filtration flow-rate was as shown in Table 3.3.

The decline in the filtration flow-rate with increase in the filtration head-loss was larger as the initial filtration flow-rate becomes large. Beyond an initial filtration flow-rate of 360 $m^3/m^2 \cdot d$, the flow-rate declined soon after the start of filtration. With reference to Fig. 3.3 in which the typical sewage flow pattern of Toba Sewage Treatment Plant is given, the flow-rate is relatively constant over a period of about 9 hrs (from 10:00 a.m. to 7:00 p.m.). From 7:00 p.m., however, the sewage hourly flow-rate to the treatment plant was on the decline. For this reason, if the actual treatment plant should be operated in declining filtration method as in the experiments (i.e., if media size and depth of filter No.2 or No.3 is applied), and if the filter should be washed once every day, the initial filtration flow-rate might have to be selected less than 360 $m^3/m^2 \cdot d$.

3.3 Effluent quality

The quality of the influent and effluent of the filter is shown in Table 3.4. The quality of filtrate was largely influenced by the quality of biological treatment effluent, making the effects of filtration flow-rate uncertain.

Filtrate quality was degraded when experiments were carried out at a filtration flow-rate of 480 $m^3/m^2 \cdot d$. This is because of the breakdown of the sewage sludge treatment facility during that period, which forced to accumulate sludge in the primary sedimentation and biological treatment facilities.

There were differences in the quality of filtrate among the three filters; filter No.1 was better than filters No.2 and No.3, and filter No.2 was slightly inferior to filter No.3. It was also found that the difference in filtrate quality among the three filters was noticeably increased when the initial filtration flow-rate was large. But, it was inferred that the difference would be very little if the filtration flow-rate was lower than 360 $m^3/m^2 \cdot d$. When three filters applying declining filtration method with media thickness set at 1.0 m and maximum filtration head at 3.0 m was carried out, the change in the filtrate quality during the filtrating period was very little. Table 3.5 shows the above two examples.

With all, it is concluded that when the filtration is to be conducted at a filtration flow-rate of less than 360 $m^3/m^2 \cdot d$, the ratio of the anthracite coal layer to the total media depth be preferably around 60 per cent, just as in the filter system configuration preferred in this filters No.2 and No.3. It is proved with this arrangement that not only can be run-length increased like the example shown in Fig. 3.1, but the filtrate quality can be maintained satisfactory.

There was no significant difference in both run-length and filtrate quality between the dual-media filter No.2 and tri-media filter No.3. This may be attributable to the selection of the size of gannet sand used as a medium.

3.4 Organic loads by washed waste water

The washed wastewater from the filter system is returned to inlet of primary sedimentation tanks of the treatment plant. In the existing sewage treatment system, the loading by this washed wastewater cannot be ignored. For this reason, the volume of washed wastewater and the loading by it were examined. Fig. 3.2 shows an example.

In case the back washing was carried out with the media expansion ratio set at 20 to 25 per cent, it took about 5 min. for the filter No.1 (with short run-length) and 7.5 to 8 min. for the filters No.2 and No.3 to attain a practical refreshness. As regards the example shown in Fig. 3.2, run-length was 13 hrs for the filter No.1, 25 hrs for the filter No.2 and 24 hrs for the filter No.3.

At the time of the back washing, the maximum BOD concentration was 939 mg/1, 1,639 mg/1 and 1,306 mg/1 for the filters No.1, No.2 and No.3, respectively.

The maximum suspended solids concentration was 1,220 mg/1, 2,320 mg/1 and 2,070 mg/1, respectively. However, during the required washing time in which excess washing time was not included (5 min. for filter No.1; 8 min. for filters No.2 and No.3), the mean BOD concentration was 287 mg/l, 410 mg/l and 411 mg/l for filters No.1, No.2 and No.3, respectively. Also, the mean suspended solids concentration was 384 mg/l, 551 mg/l and 575 mg/l, respectively. While the maximum BOD concentration and maximum suspended solids concentration were so high as above, the mean values were about 2 to 3 times those of the primary influent. The washing water required was 5.9 to 7.6 m^3/filter m^2, and retio of washing water to filtrate volume was less than 2.5 per cent.

3.5 Influences of filter washed wastewater on primary sedimentation tank of existing facilities

Investigations were conducted on the influences of washed wastewater created by once-a-day washing of the filter over the functions of the primary sedimentation tank of the existing facilities into which it would be return. Fig. 3.3 is an example showing the influent load to the primary sedimentation tank of Toba Sewage Treatment Plant. During the survey, the sewage influent at the Plant was 362 x 10^3 m^3/d.

The Toba Plant handles the sludge of the nearby Kisshoin Sewage Treatment Plant and treats supernatant coming from another anaerobic digestion facility which was treating night soil. As a consequence, the return waste water from the plant at that time was as much as 41 x 10^3 m^3/d running into the primary sedimentation tanks. The over-all influent of the primary sedimentation tanks was 403 x 10^3 m^3/d, and the mean surface loading of the tanks was 22.7 $m^3/m^2 \cdot d$. The average removals of suspended solids and BOD were 76.5 and 59.2 per cent respectively. As given in Fig. 3.3, the influent load to the primary sedimentation tanks was more than 20 x 10^3 m^3/hr during a 10-hr daytime, declining gradually from 7:00 p.m. to 6:00 a.m., and thence increased sharply during the period from 6:00 a.m. to 10:00 a.m.

Both the influent suspended solids load and BOD load showed a tendency to decrease from midnight to early morning.

Assuming that at Toba Plant a dual-media or tri-media filter system is installed and operated at a filtration flow-rate of 300 $m^3/m^2 \cdot d$ with washing once a day, the number of filters required will be a dozen or so. If the filters are to be recleaned one by one sequentially, the washed wastewater load on the primary sedimentation tank will be equalized over some period of time. The hatched part appearing in Fig. 3.3 is an example showing the case where the filters are washed during 2:00 a.m. to 8:00 a.m. period. In this example, the overflow rate of the primary sedimentation tanks will be only 21.4 $m^3/m^2 \cdot d$ even at 2:00 a.m. when the situation is most critical. This is almost on the same level as the mean overflow rate of 22.8 $m^3/m^2 \cdot d$ of the primary sedimentation tanks without filters.

BOD loading and suspended solids loading are not more than 1/5 the mean values when filters are not installed. Namely, the filter washing, if taken from midnight to early morning, would not affect the existing sewage treatment facilities and could do without expansion of the existing facilities.

Feasibility of treating the washed wastewater in the primary sedimentation tank was also examined using a settling column. The results revealed that suspended solids in the washed wastewater could satisfactorily settle out and be removed even without dosage of a catonic polymer as the flocculant aid. (see Fig. 3.2)

3.6 Investigations now in progress

(1) As described in the foregoing, comparative studies on the three down-flow filters have been carried out so far. It was reported that in England the up-flow filtration has been sucessfully used for polishing of the secondary effluent.
To follow suit, one Body Immedium Filter (filtration area: 1.2 m^2) was installed at this plant, and has been studied in comparison with other three media filters.

The media (silica sand) used for the up-flow filter are in two layers - one having an effective size of 1.16 mm and a uniformity coefficient of 1.33 and another having 2.01 mm and 1.31, respectively. The over-all media depth of the filter is 1.55 m. The up-flow filter is cleaned by air agitation and back washing. For the filter washing water, the secondary effluent is used.

As shown in Table 3.6, the filtrate quality obtained is little different from that by the down-flow filters No.2 and No.3. But, the up-flow filter requires twice or more time in washing than the down-flow filter does. However, the up-flow filter, permits to used the secondary effluent directly as the filter washing water, and to dispose the effluent without post aeration because dissolved oxygen concentration in the filtrate is relatively high.

(2) At the primary and secondary processes of large-scale sewage treatment plants in Japan, where wastewater flows in from combined sewarage systems, phosphorus is removed more than 60% from influent of primary sedimentation tank in which digested supernatant is contained with raw sewage. In some cases, removal of phosphorus in the effluent is attained more than 80%. However, stringent effluent standard for phosphorus may be set for sewage treatment plant effluent, and additional removal of phosphorus may be required. It will therefore be of great necessity to find an economical way for the reduction of residual phosphorus in secondary effluent.

Certain data from the United States indicate that the concentration of phosphorus in secondary effluent is 6 mg/l or higher in the U.S., while that in Japan is 2 mg/l or lower at large-scale sewage treatment plant at present. Therefore, less removal of phosphorus from secondary effluent might be allowed in Japan.

Investigations are in progress as to the influences of flush mixing with coagulant addition as a pretreatment of filtration over the filtrability, and quality of filtered water and washed wastewater. According to a laboratory test, it is found that the method improves removal of phosphates and COD_{Mn}. The mole ratio of Al^{+3} to the concentration of phosphorus in secondary effluent, however, is required to be 2.5 to 3 for maintaining the concentration in filter effluent lower than 0.5 mg/l, and the economical aspect of phosphorus removal should be examined carefully.

(3) Since the reduction of COD_{Mn} is not enough when only direct filtration of secondary effluent is applied for an additional process, experiments on granular activated carbon adsorption process have been started. The granular activated carbon adsorption system used is composed of six down-flow gravity type pilot contactors each having an adsorption area of 0.7 m^2. The system is designed to be modified into three running modes - 1-tank 6-parallel, 2-tank 3-parallel and 3-tank 2-parallel - whichever is desired. In the experiments by 2-tank 3-parallel or 3-tank 2-parallel, the so-called merry-go-round operation is possible.

The season why the granular activated carbon adsorption tank is made of the down-flow gravity type is that when this kind of facility is installed in the future its main body is expected to be constructed in reinforced concrete structure.

The effluent of the activated carbon adsorption contactors is continuously measured with a 30° V-notch weir equipped with an automatic level gauge.

For the measurement and sampling of washed waste water, a 90° V-notch weir box is installed. The experimental facility can be operated semi-automatically or full automatically or manually. At present, 2-tank 3-parallel operating mode is employed, and granular activated carbon of Calgon 8 x 30 meshes, Calgon 12 x 40 meshes and Shirasagi 8 x 30 meshes is tested. Carbon bed depth is 3.0 m for all of the contactors..

Influent is 240 $m^3/m^2 \cdot d$, and back washing of carbon bed is carried out once a day. An example of the water quality data is shown in Table 3.7.

3.7 Aspects of future studies

At the pilot plant located at Toba Sewage Treatment Plant, Kyoto, a furnace for regeneration of exhausted granular activated carbon will be installed in order to test the replicate use of regenerated activated carbon. Also, up-flow type granular activated carbon adsorption contactors will be installed for the comparative study with the down-flow type granular activated carbon adsorption system.

As regards to the ammonia removal, laboratory tests have already been pushed forward for the installation of a break-point chlorination pilot facility. It is in the stage of design of the instrumentation. In Japan, water pollution problems due to excess nitrogen in river and lake water and irrigation water have come to stay, and the development of an economical way of removing nitrogen has been voiced for.

Unfortunately, however, each sewage treatment plant barely possess space to enlarge the facilities, and has only a limited space to accommodate such extra work. With this in mind, experiments of removing nitrogen are started [illegible] the break-point chlorination process for which space can be small.

The soda industry here in Japan has been urged to change its production process from mercury process to membrane one and it remains uncertain whether so much chlorine gas will be avilable for the break-point chlorination process from now on. Therefore, it is necessary to investigate nitrogen removal processes more extensively.

Another facility to be installed will be ion-exchange processes for the removal of nitrate and phosphate, and also ammonia.

Table 3.1 Media Size Using Experiment

Media	Effective Size (mm)	Uniformity Coeficient
Anthercite Coal	1.22	1.21
Silica Sand	0.87	1.21
Garnnet Sand	0.54	1.41

Table 3.2 Media Depth of Each Pilot Filter

Media \ Filter	No.1 (mm)	No.2 (mm)	No.3 (mm)
Anthercite Coal	150	625	625
Silica Sand	850	375	300
Garnnet Sand	-	-	75
Total	1,000	1,000	1,000

Fig. 3.1 Relationship Between Initial Flow-rate and Average Run-Length of Pilot Filters

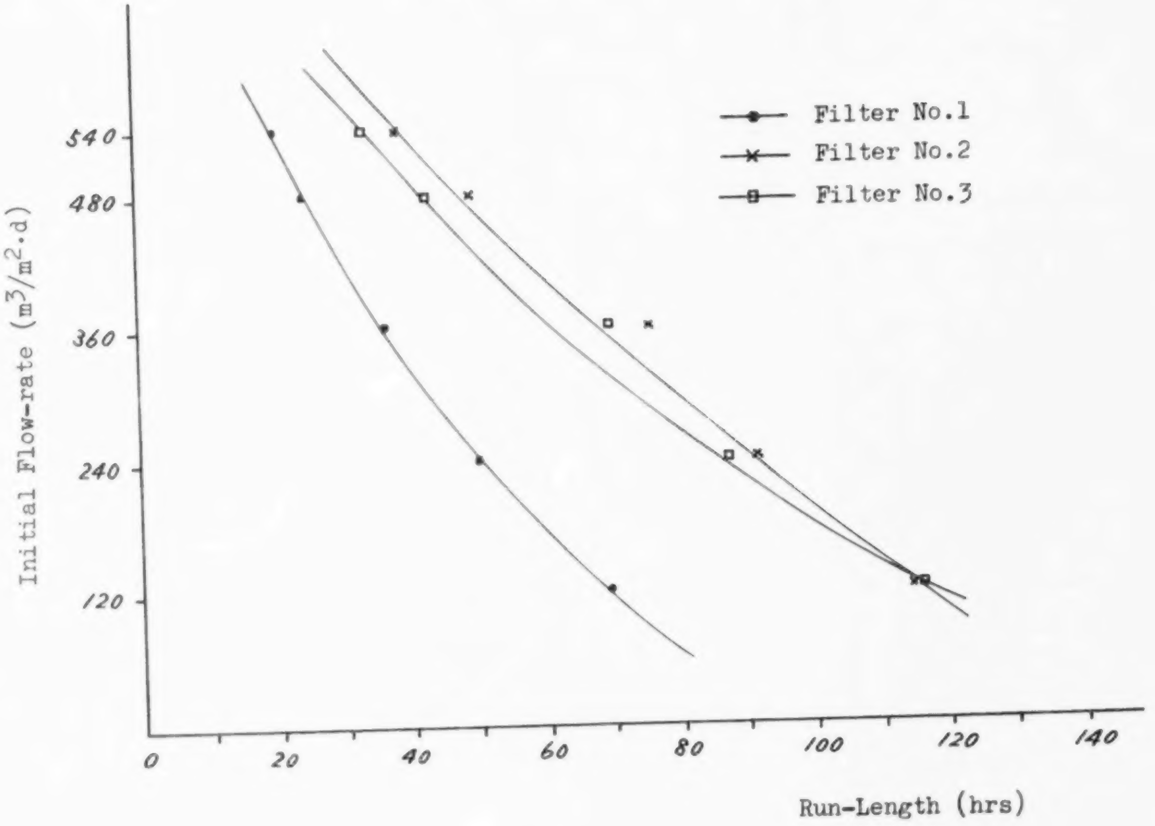


Table 3.3 Relationship between Filtration Flow-rate Declined and Filtration Head-Loss

($m^3/m^2 \cdot d$)

Initial Flow-rate ($m^3/m^2.d$)	Filtration Head-Loss (cm)					
	50	100	150	200	250	300
120	118	108	98	85	78	58
240	232	218	200	182	158	124
360	358	325	307	278	236	197
480	480	462	410	365	322	258
540	540	500	472	405	350	292

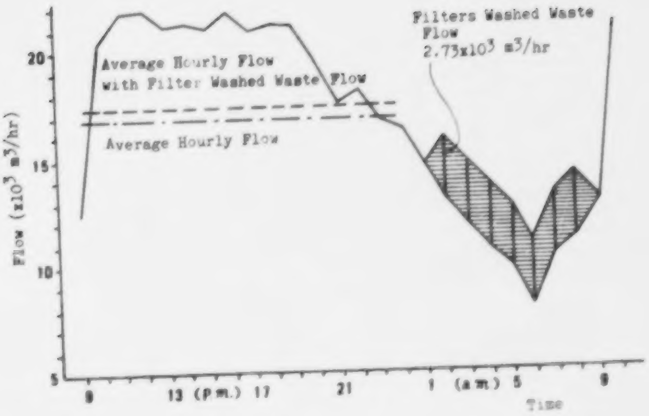
Table 3.4 Influent and Filtrate Quality of Pilot-Filters

Initial Filtration Flow-rate ($m^3/m^2 \cdot d$)		120				240				360				480				540			
Items / Influent or Filtrate		Inf.	Filtrate			Inf.	Filtrate			Inf.	Filtrate			Inf.	Filtrate			Inf.	Filtrate		
			No.1	No.2	No.3		No.1	No.2	No.3		No.1	No.2	No.3		No.1	No.2	No.3		No.1	No.2	No.3
Nos. of Data		8	5	8	8	9	8	9	9	10	7	10	10	10	7	10	10	9	7	9	9
S.S. (mg/l)	max.	26.3	8.3	8.1	7.4	10.8	3.7	3.8	3.1	28.8	2.4	3.2	3.0	67.7	29.0	43.5	46.5	33.5	5.6	19.0	18.3
	min.	7.9	1.0	1.2	1.6	4.6	0.5	1.0	1.0	3.1	0.3	0.2	0.1	5.0	0.5	0.8	0.7	4.2	1.4	1.2	1.1
	Av.	14.8	3.7	4.0	3.9	7.5	2.2	2.7	2.5	9.7	1.3	1.4	1.3	22.8	6.6	11.4	12.0	13.8	3.5	5.4	5.2
BOD (mg/l)	max.	29.1	10.9	11.2	9.9	17.9	4.0	7.3	6.8	16.9	4.1	3.6	3.4	32.0	7.5	19.4	21.0	29.9	7.1	20.2	16.1
	min.	12.5	1.7	2.1	2.2	6.0	1.9	1.8	2.2	2.5	1.0	1.4	0.6	4.4	1.0	1.4	1.3	6.3	1.5	1.3	1.0
	Av.	22.4	3.9	4.1	3.8	10.9	3.2	4.3	4.3	10.8	2.1	2.2	2.0	18.0	2.9	6.4	6.3	15.8	4.1	5.6	4.8
COD_{Mn} (mg/l)	max.	20.7	14.3	15.1	14.9	15.7	12.8	13.7	13.0	23.8	16.4	16.4	16.7	28.6	17.9	20.6	21.5	26.0	16.0	21.2	21.7
	min.	9.8	8.4	8.4	8.4	5.4	1.0	1.0	1.0	6.1	4.5	4.9	4.8	8.8	5.5	5.9	5.4	9.0	8.2	8.1	8.4
	Av.	15.6	10.4	11.4	11.3	12.2	9.8	10.2	9.8	12.5	10.4	10.1	9.8	16.0	9.9	12.0	11.9	16.5	12.6	12.8	13.0

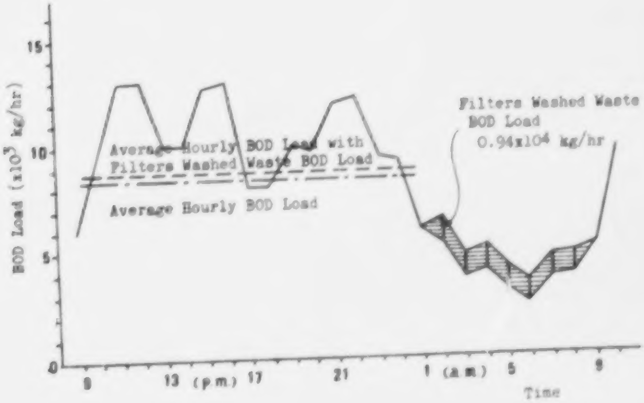
Table 3.5 Comparision to Filtrate of Three Filters after Filtration Start

How-rate $(m^3/m^2\cdot d)$	Influent or Filtrate / Items		Inf.	Filtrate			Inf.	Filtrate			Inf.	Filtrate		
				No.1	No.2	No.3		No.1	No.2	No.3		No.1	No.2	No.3
240	Sampling Time after Filtration Start (hrs:min)		2:45		3:00		26:25		27:00		46:25		47:00	
	Liquid Temp (°C)		24.6	25.5	25.5	25.5	24.8	25.2	25.2	25.2	24.5	25.0	25.0	25.0
	Dissolved Oxygen (mg/l)		0.5	0.5	0.3	0.3	0.2	0.2	0.2	0.2	0.4	0.2	0.2	0.2
	S.S.	conc. (mg/l)	8.0	3.7	2.9	3.1	10.7	2.6	3.3	2.7	6.4	2.8	2.9	2.8
		rem. (%)	-	53.7	63.8	61.2	-	75.7	69.1	74.7	-	56.6	54.8	56.3
	BOD	conc. (mg/l)	6.2	3.5	3.1	3.3	12.9	3.4	3.4	3.2	12.5	4.0	3.8	3.2
		rem. (%)	-	43.6	50.0	46.8	-	73.5	73.5	75.1	-	68.0	69.5	74.5
	COD_{Mn}	conc. (mg/l)	10.4	9.2	10.1	10.4	12.4	9.3	9.7	9.7	12.6	11.2	11.0	9.5
		rem. (%)	-	11.5	28.8	0.0	-	25.0	21.8	21.8	-	11.1	12.7	24.6
480	Sampling Time after Filtration Start (hrs:min)		2:43	3:00			24:43	25:00						
	Liquid Temp. (°C)		26.2	26.8	26.8	26.8	26.0	26.5	26.5	26.5				
	Dissolved Oxygen (mg/l)		0.2	0.2	0.2	0.2	1.7	0.2	0.1	0.1				
	S.S.	conc. (mg/l)	8.0	1.8	1.7	1.7	8.2	2.0	1.5	1.7				
		rem. (%)	-	77.5	78.8	78.8	-	75.6	81.7	79.4				
	BOD	conc. (mg/l)	6.2	2.0	1.6	1.6	4.4	1.4	1.4	1.7				
		rem. (%)	-	67.7	74.1	74.1	-	68.2	68.2	61.5				
	COD_{Mn}	conc. (mg/l)	9.8	8.4	8.5	8.1	9.6	9.0	9.0	9.0				
		rem. (%)	-	14.3	13.3	17.4	-	6.3	6.3	6.3				

(a) Sewage Flow



(b) BOD Load



(c) Influent Suspended Solids Load

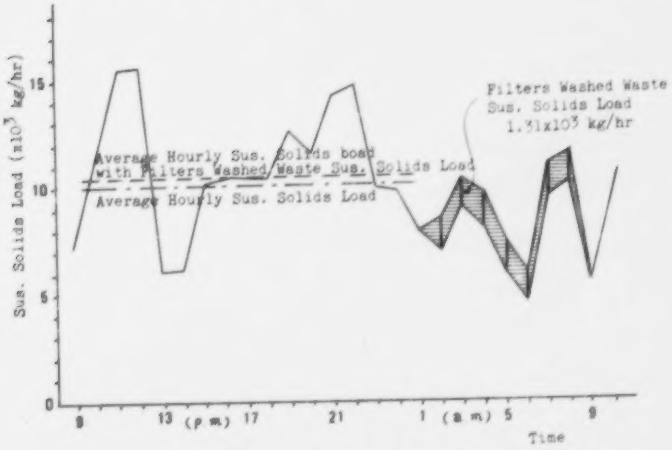


Fig. 3.3 Typical Flow and Load Pattern to Primary Sedimentation Tanks Assuming to be Equipped Filters, Toba Sewage Treatment Plant, Kyoto

Fig. 3.2 Settling Column Tests of Filter Washed Wastewater

Column height 2.2 m
Column Diameter 0.3 m

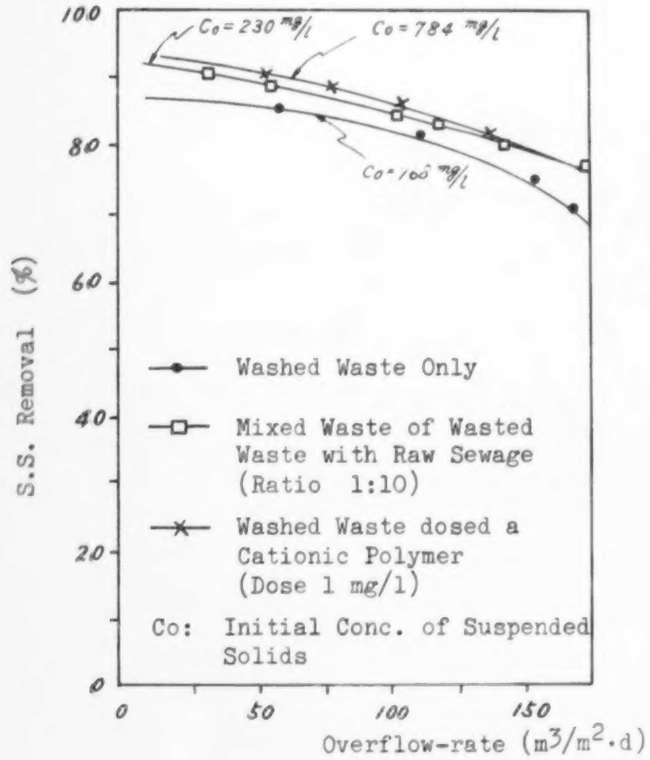


Table 3.6 Comparision to Filtrate both Down-flow Filters and Up-flow Filter

Items \ Influent or Filtrate		Influent	Filtrate			
			No.1	No.2	No.3	No.4
Liquid Temp. (°C)		20.3	20.7	20.7	20.7	20.7
Dissolved Oxygen (mg/l)		0.5	0.5	0.5	0.5	6.2
Turbidity (mg/l)		7.5	2.65	2.55	3.60	3.20
S.S.	conc. (mg/l)	11.2	2.7	3.3	4.9	3.9
	red. (%)	-	75.8	71.3	56.2	65.0
BOD	conc. (mg/l)	17.8	5.5	7.1	9.2	8.2
	red. (%)	-	69.2	60.1	45.5	54.0
COD_{Mn}	conc. (mg/l)	13.7	11.7	10.9	11.7	11.9
	red. (%)	-	14.6	20.4	14.6	13.1
TOC	conc. (mg/l)	14.2	9.6	11.2	10.0	10.8
	red. (%)	-	32.4	21.1	29.6	23.9

Remark: Filtration Flow-rate : 240 $m^3/m^2 \cdot d$

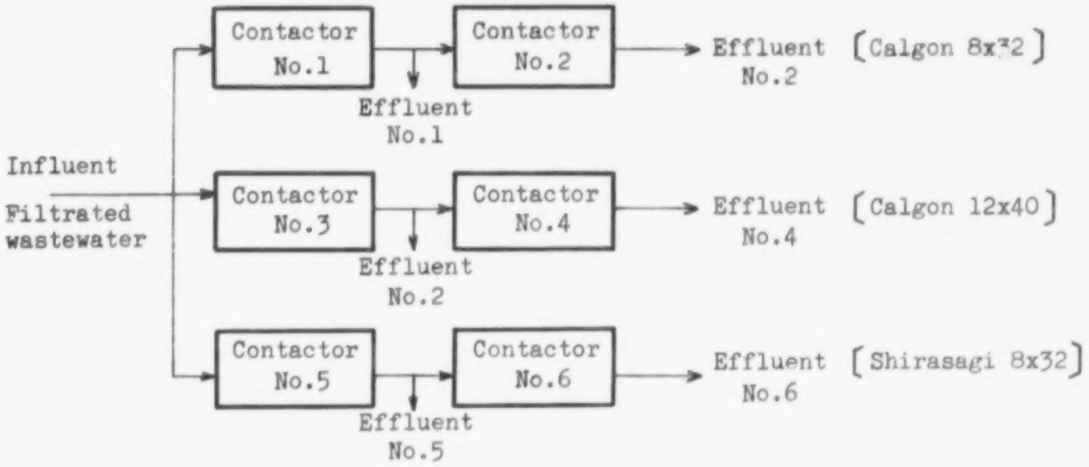
Down-flow Filter : No.1, No.2, No.3
(refer to Table 3.1 & 3.2)

Up-flow Filter : No.4

Table 3.7 Influent and Effluent Quality of Pilot Granular Activated Carbon Adsorption Contactors

Items \ Influent or Effluent		In-fluent	Effluent					
			No.1	No.2	No.3	No.4	No.5	No.6
Liquid Temp. (°C)		18.0	17.4	17.4	17.4	17.4	17.4	17.4
Dissolved Oxygen (mg/l)		9.3	0.4	1.0	1.2	0.7	0.7	0.6
S.S.	conc. (mg/l)	3.0	3.5	1.9	2.8	2.3	3.6	2.4
	rem. (%)	-	-	36.7	6.7	23.3	-	20.0
BOD	conc. (mg/l)	5.3	5.1	5.0	5.1	3.8	4.6	4.3
	rem. (%)	-	3.8	5.7	3.8	28.3	13.2	18.9
COD_{Mn}	conc. (mg/l)	11.3	7.7	4.6	6.6	3.5	7.2	4.8
	rem. (%)	-	39.7	59.3	41.6	69.0	36.3	57.5
Kjeldahl Nitrogen	conc. (mg/l)	6.3	3.8	3.2	4.8	2.4	4.2	2.4
	rem. (%)	-	39.7	49.3	23.8	61.9	33.3	61.9

Remark: Arrangement of Granular Activated Carbon Adsorption Contactors



Adsorption Time in Each Contactor : 15 min

4. Evaluation of treatability depending upon water quality matrices

Connecting with the treatability of waste water, the status of the impurities could be represented by three components such as the impurity size, chemical property and cencentration (1), (2), (3). Thus if the impurity sizes are devided into several classes by means of mechanical, optical or gel chromatographical technique, a water quality matrix will be able to be established schematically as shown in Figure 4.1.

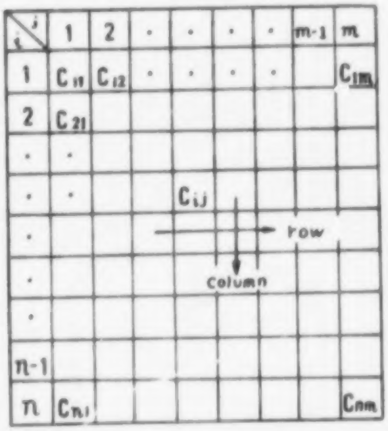


Fig. 4.1 Water Quality Matrix

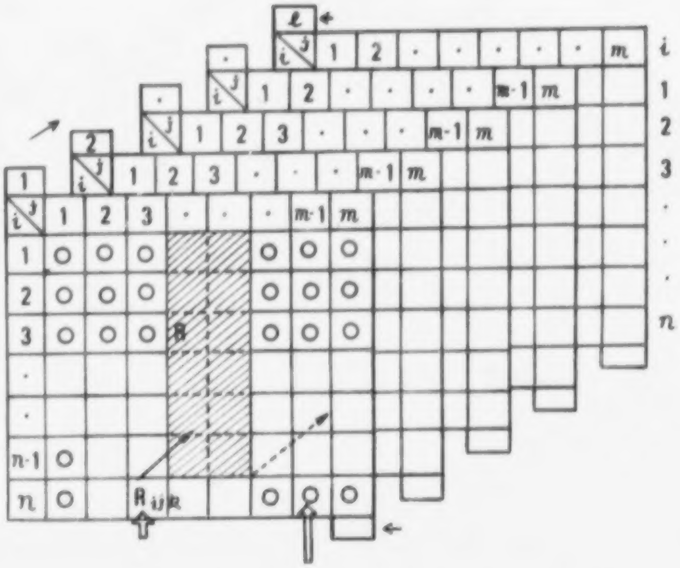


Fig. 4.2 Water Quality Conversion Matrices

The water quality conversion matrices as shown in Figure 4.2 are prepared for each process of water treatment. The indices of the columns and rows are the same with the water quality matrix, but the elements are written as the percentage of removal, R(ij), instead of the concentration, C(ij), of the water quality matrixes. If the concentration is greatly different from each other, different treatment process shall be proposed or the different percentages of removal shall be put into the matrix elements. Thus, water quality conversion matrices should be shown as the three dimensional matrices to take the concentration effects into account.

In these water quality conversion matrices, while no removal is anticipated, the element R(ijk) is zero., and for the removed part the element R(ijk) takes a positive value. In one case when some new components such as a metabolic waste of biological treatments are added by the treatment, the elements become negative. In another case when no water quality elements exist prior to the treatment at the components, basic elements to be taken into the calculation should be pointed out. In addition, if the element R(ijk) cannot be decided independedtly, the relationship between other connected components should be described.

In order to develope comprehensive technology for water quality management on the basis of an idea as mentioned above, the research project supported by Special Fund for Promoting of Multiministerial Research Project under the Jurisdiction of the Science and Technology Agency has been carried on since 1972 to 1974. The project director is prof. Kenji Tambo, Hokkaido University and the grantees are The Institute of Public Health, Ministry of Health and Welfare and Kinki Regional Construction Bureau, Ministry of Construction.

4.1 Experiments for the verification of the concept

A series of laboratory experiments and pilot plant operations were carried out to evaluate the feasibility of the above mentioned concept. To obtain general results without falling into local conditions, many kinds of raw waters and treated waters in widely spread areas of Japan were used for the experimental works.

1) Laboratory scale coagulation and carbon adsorption experiments

The major purpose of those laboratory scale experiments was to study performance of the activated sludge process and chemical coagulation process together with the results of followed activated carbon adsorption process. For the purpose, the following experimental studies were carried out.

1. Treatment of raw sewage by combination of alum or lime coagulation, sedimentation, filtration, and activated carbon adsorption.
2. Treatment of the activated sludge process plant (Makomanai, Sapporo) effluent by combination of alum coagulation, sand filtration, and activated carbon adsorption.
3. Treatment of the activated sludge process plant (Fusiko, Sapporo) effluent by combination of a simple sand filtration without coagulation and activated carbon adsorption.

The laboratory scale semi-continuous treatment apparatus as shown in Figure 4.3 consisted of a 200 liter of settling tank with coagulation facility, a rapid sand filter, a filtered water storage, and four down flow granular activated carbon columns in series.

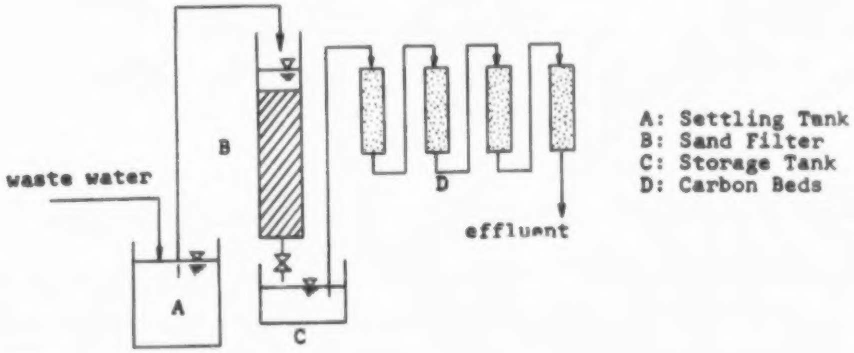


Fig. 4.3 Flow diagram of laboratory scale apparatus

Standard jar tests were used to determine the coagulant dosage. A coagulant was added to the settling tank, then, rapidly mixed for 5 minutes, flocculated for 30 minutes, and settled for 1 hour. The supernatant from the tank was applied to the sand filter with 50 cm of sand bed at constant rate of 50 m/day. The filtered water was once put into the storage tank, then fed to the four activated carbon columns arranged in series. The size of the carbon column was 3 cm in diameter and 40 cm in length. Pittsburg CAL activated carbon was placed in those columns with 25 cm depth. The rate of the flow through the adsorption columns was set as 144 m/day.

In parallel with the laboratory scale continuous adsorption experiments, batch adsorption tests were also carried out by using powdered activated carbon.

The turbidity was measured by a photo-turbidimeter which was calibrated with standard kaolinite solution. The total organic carbon (TOC) concentration was determined by a Toshiba-Beckman Carbonaceous Analyzer. Sample water having organic carbon concentrations below 1 to 2 mg/l was concentrated under reduced pressure at the normal temperature prior to the TOC determination. Sometimes, the pH of the water samples was adjusted at pH=2 and the samples were shaken or aerated with N_2 gas so as to remove inorganic carbon which was apt to hinder the TOC determination. Ultraviolet absorbances were determined by a Hitachi 124 double beam spectrophotometer. The concentrations of protein (Folin-Chiocalteu test), amino acids (Ninhydrin test), and carbohydrates (Anthrone test) were also determined for the same samples. The COD and major inorganic constituents were also determined after the standard methods (4).

2) Physico-chemical treatment of sulphide pulp waste water

A physico-chemical treatment of sulphide pulp waste water from a pulp mill in Tomakomai, Hokkaido, by using activated carbon produced from the lignin of the pulp waste water, was carried out to verify the applicability of the concept to the outside of the municipal waste water treatment.

The flow sheet of the sulphide pulp waste water treatment processes and the lignin base activated carbon production from the waste water is shown in Figure 4.4.

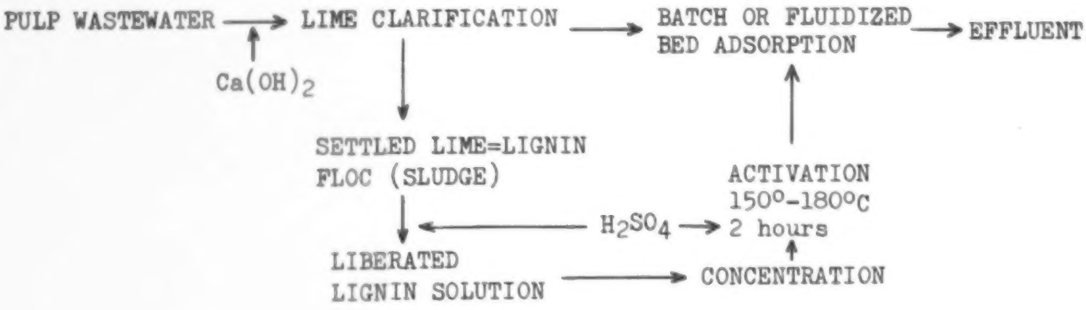

PULP WASTEWATER
LIME CLARIFICATION
BATCH OR FLUIDIZED BED ADSORPTION
EFFLUENT
Ca(OH)2
SETTLED LIME=LIGNIN FLOC (SLUDGE)
ACTIVATION 150°-180°C 2 hours
H2SO4
LIBERATED LIGNIN SOLUTION
CONCENTRATION

Fig. 4.4 Flow sheet of the pulp wastewater treatment and the production of powdered activated carbon from lignin of the waste water.

3) Pilot plant operations

To confirm the results of the laboratory scale experiments and solidify the concept, two series of the same type physico-chemical pilot plants were constructed and operated at Soseigawa sewage treatment plant in Sapporo. A series of the plants treated raw sewage which was pumped up from the primary settling tank overflow of the Soseigawa sewage treatment plant. Another series of the pilot plant treated the activated sludge process effluent of the sewage treatment plant for the purpose of the tertiary physico-chemical treatment processes with or without chemical coagulation process.

A schematic flow diagram of a series of the pilot plant processes is shown in Figure 4.5. The operational flow chart including operation conditions is written in Figure 4.6.

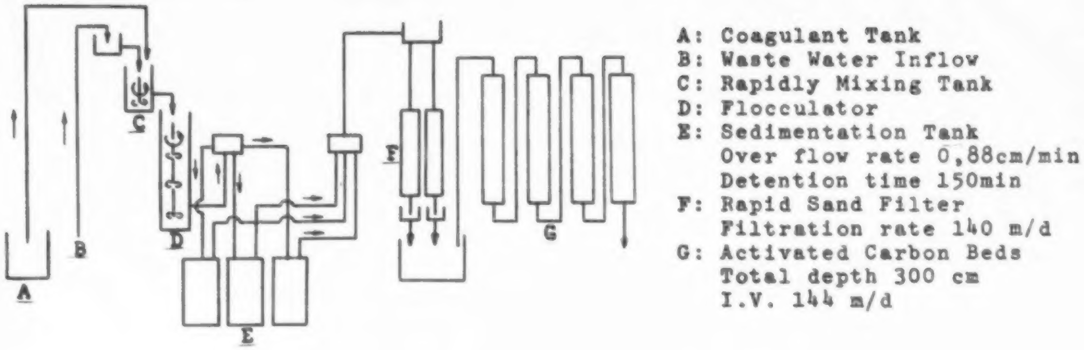


Fig. 4.5 Flow diagram of the pilot plant

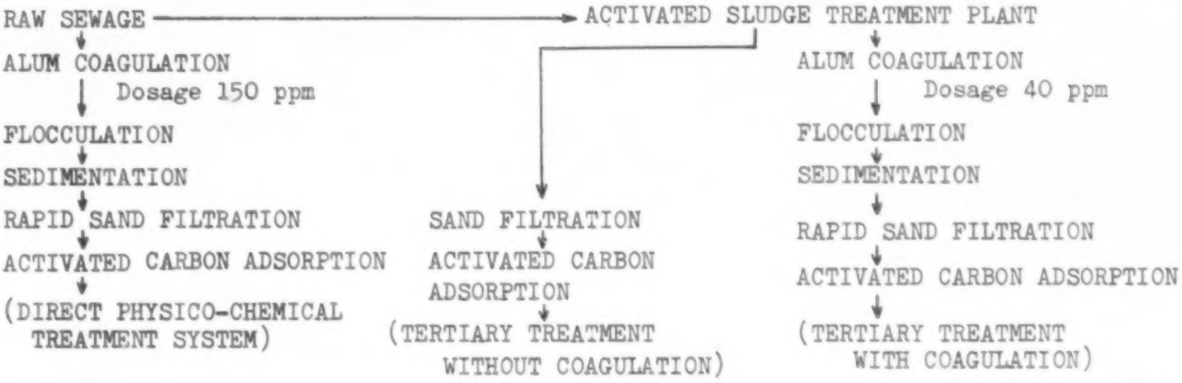

RAW SEWAGE
ALUM COAGULATION
Dosage 150 ppm
FLOCCULATION
SEDIMENTATION
RAPID SAND FILTRATION
ACTIVATED CARBON ADSORPTION
(DIRECT PHYSICO-CHEMICAL TREATMENT SYSTEM)
ACTIVATED SLUDGE TREATMENT PLANT
SAND FILTRATION
ACTIVATED CARBON ADSORPTION
(TERTIARY TREATMENT WITHOUT COAGULATION)
ALUM COAGULATION
Dosage 40 ppm
FLOCCULATION
SEDIMENTATION
RAPID SAND FILTRATION
ACTIVATED CARBON ADSORPTION
(TERTIARY TREATMENT WITH COAGULATION)

Fig. 4.6 Operational flow chart of the pilot plants.

4) Gel chromatogram

The impurity size distribution of minute fractions of those samples which had passed through 0.45 micron meter membrane filters was determined by a gel filtration mainly on a column of Sephadex G-15. For some samples, Sephadex G-25 or G-50 was also used. Those gels were swelled for 24 hours prior to setting into the columns. The resulting slurry of the gel was then poured into the glass columns of 2.5 to 4 cm in diameter and 100 cm in length with 90 cm of the gel bed height.

Most water samples were once filtered by 0.45 micron meter membrane filter and concentrated into suitable concentrations by rotary vacuum evaporators at the temperatures of 25 to 30°C before applying to the column. Concentrated samples of 10 to 30 ml were placed at the top of the gel beds. Distilled water was introduced into the column from the top at a constant rate of flow for the elution. The rate of flow was 20 to 50 ml/min. Effluent from the column bottom was collected in the test tubes by an automatic fraction collector as 10 to 20 ml of successive fractions. For each fraction, measurements were carried out for TOC content and 220 to 280 millimicron meters of ultraviolet absorbance.

The impurity size distribution characteristics of the gel chromatogram were calibrated from the elution volume of Blue Dextran (M.W. 2 millions), raffinose (M.W. 504), maltose (M.W. 342), glucose (M.W. 180), ethylene glycol (M.W. 62), methanol (M.W. 32), and DBS (M.W. 349).

4.2 Results of Experiments

1) Results of laboratory scale coagulation and carbon adsorption experiments

The effectiveness of alum coagulation followed by carbon adsorption is appreciated by comparing Tables 4.2 and 4.2. Some amount of leakage of $KMnO_4$ COD and E250 (Absorbance at 250 millimicron meters) constituents are recognized when the coagulation process is omitted. Whereas, in the case when alum coagulation process is included, the leakage from the carbon bed is very scarce in spite of a relatively higher raw water concentration.

Table 4.1 Non coagulated, filtered and carbon adsorbed secondary effluent.

Hours of operation	$COD(KMnO_4)$ ppm		E250	
	Filtered	Adsorbed	Filtered	Adsorbed
48	4.5	1.0	0.122	0.024
96	5.0	1.0	0.109	0.018
168	4.5	0.9	0.110	0.020
336	5.0	1.2	0.137	0.022

Table 4.2 Alum coagulated, filtered and carbon adsorbed secondary effluent.

Hours of operation	$COD(KMnO_4)$ ppm		E250	
	Filtered	Adsorbed	Filtered	Adsorbed
40	8.2	0.3	0.094	0.000
88	8.5	0.2	0.137	0.006
160	8.3	0.9	0.100	0.006
358	8.0	0.9	0.098	0.002

The gel chromatographic studies shown in Figures 4.7 and 4.8 reveal that without coagulation there exists typical leakage of larger size impurities in the Zones 1 and 2 in the carbon adsorption bed effluent. Inefficiency of activated carbon adsorption in removing larger size impurities should be marked.

A result of physico-chemical raw domestic sewage treatment (coagulation, filtration, and carbon adsorption) without biological process is shown in Figure 4.9. There is the leakage of COD from the very early stage of the run. The major constituents of the leaked COD are lower molecular weight substances such as amino acids, low molecular fatty acids, and carbohydrates which could be rather easily removed by biological processes.

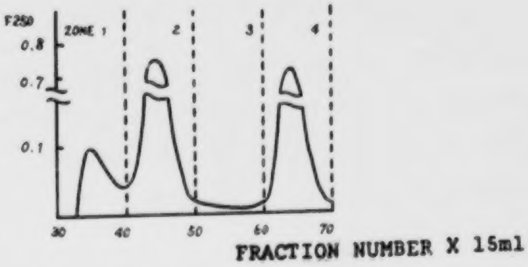


Fig. 4.7 Gel chromatogram of secondary effluent

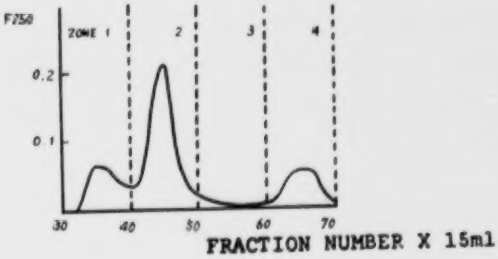
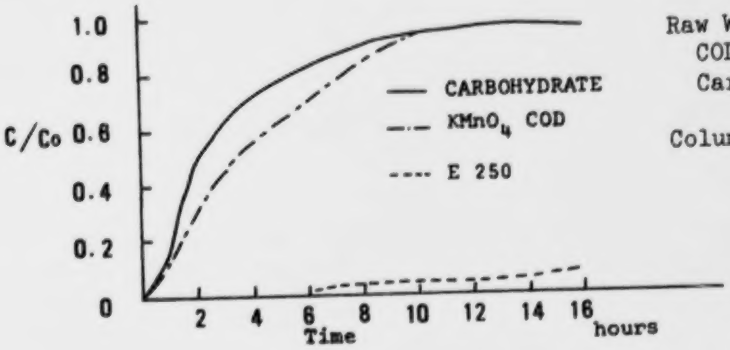


Fig. 4.8 Gel chromatogram of carbon adsorbed secondary effluent without coagulation



Raw Water Conc.
COD 122 ppm
Carbohydrate 152 ppm

Column depth 20 cm

Fig. 4.9 Physico-chemical treatment effluent of domestic raw sewage.

2) Results of sulphide pulp waste water treatment studies

For the physico-chemical treatment of sulphide pulp waste by using lime or alum coagulation and the laboratory-made lignin waste activated carbon adsorption, the usefulness of the above mentioned chromatographic characteristics are verified as follows.

Ineffectiveness of activated carbon adsorption and effectiveness of alum or lime coagulation in removing larger molecular substances are also found by the results being shown in Figures 4.10 and 4.11. The treatment system connecting both of the lime coagulation and fluidized carbon adsorption bed in series gives a satisfactory removal of the lignin content from the pulp waste water. The spent lignin carbon saturated with the wastewater contents is possible to be regenerated by concentrated sulphuric acid at the temperatures of 150 **to** 180°C for the reaction period of two hours.

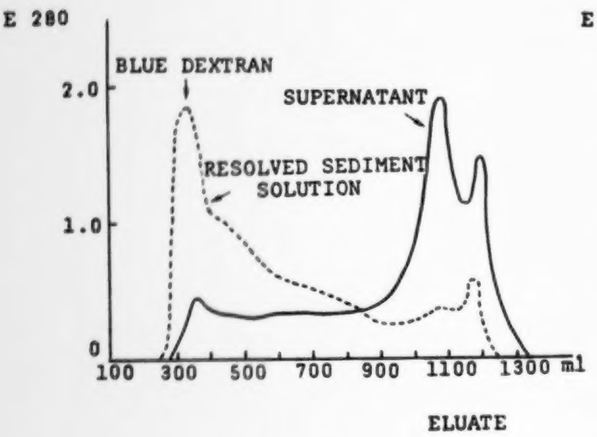


Fig. 4.10 Gel chromatograms of supernatant and resolved sediment solution of lime coagulation. Sephadex G-50

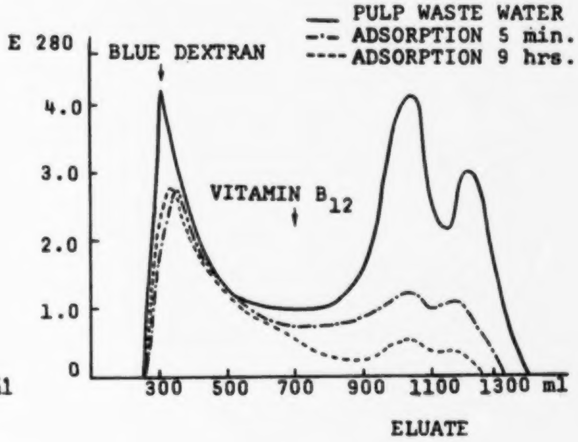


Fig. 4.11 Gel Chromatograms of pulp wastewater and carbon adsorbed wastewater. Sephadex G-50

3) Result of pilot plant studies

Typical gel chromatograms obtained from the pilot plant studies are shown in Figures 4.12, 4.13, 4.14, 4.15, 4.16 and 4.17.

As is seen in Figure 4.12, so called soluble impurities in the municipal wastewater could be divided into five groups by the Sephadex G-15 gel chromatography by using TOC, E260, and E220 as the water quality indices. This grouping is recognized as a popular one by many analyses of the samples obtained from many sewage treatment plants and natural rivers and lakes being polluted by municipal wastewaters.

Comparison of Figure 4.12 with Figure 4.13 shows that activated sludge treatment processes can effectively remove only a portion of TOC which is insensitive to the ultraviolet absorbance at 260 millimicron meters, i.e., E260. Thus, for the removal of TOC contents corresponding to E260, activated sludge processes are not effective. The increase of ultraviolet absorbance at 220 millimicron meters, i.e., E220, in the Zone 4 would be caused by some metabolic wastes of the biological activities.

Chemical coagulation followed by sedimentation and sand filtration is effective for the removal of impurities in the Zone 1 and less effective for smaller impurities in the Zones 2 to 5 as clearly seen by the comparisons of Figure 4.13 with 4.14 and Figure 4.12 with 4.16.

However, a noticeable amount of TOC removal in the Zone 3 is detected through the direct physico-chemical treatment process as shown in Figures 4.16 and 4.17. This phenomenon might be caused by some biological actions in the sand bed and carbon adsorption bed. Incidentally, no removal in the Zone 3 is observed in case of batch treatments using paper filters and powdered activated carbon or short period experiments performed by small scale laboratory tests. The total TOC contents removed by the activated carbon column in the pilot plant are far beyond the saturation adsorption capacity measured by batch tests.

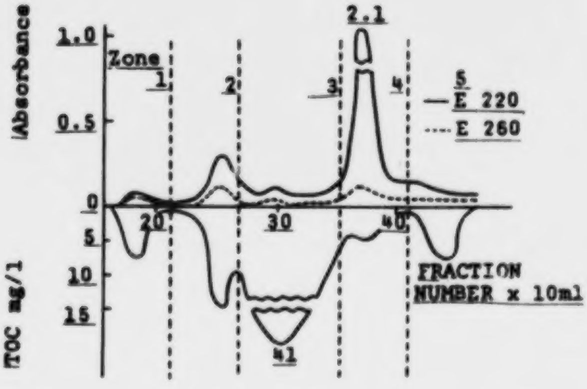


Fig. 4.12 Gel chromatogram of raw sewage

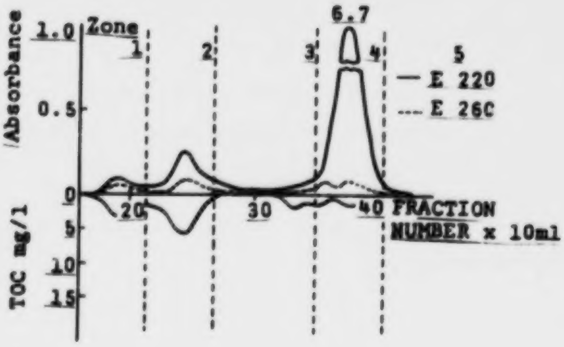


Fig. 4.13 Gel chromatogram of secondary effluent

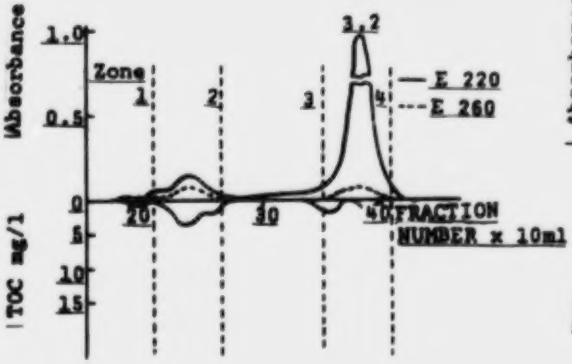


Fig. 4.14 Gel chromatogram of coagulated secondary effluent

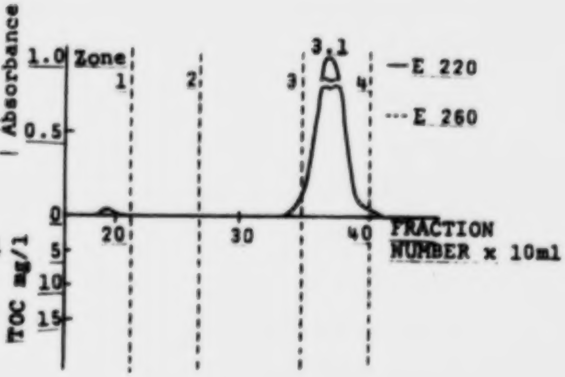


Fig. 4.15 Gel chromatogram of coagulated and adsorbed secondary effluent

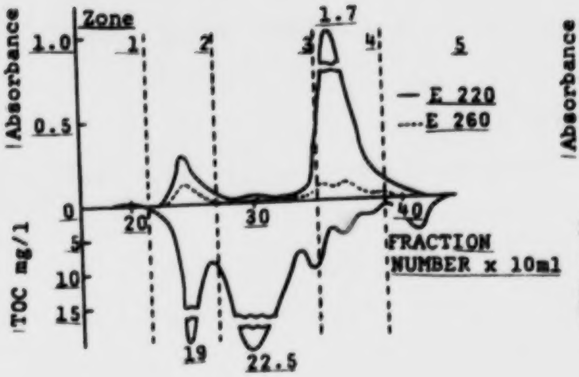


Fig. 4.16 Gel chromatogram of coagulated raw sewage

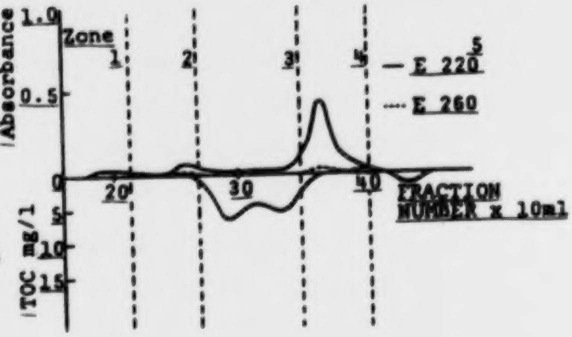


Fig. 4.17 Gel chromatogram of coagulated and adsorbed raw sewage

The TOC constituents in the Zone 3 are mainly low molecular carbohydrates, amino acids, and low molecular fatty acids that are easily decomposed by biological actions.

By the combinations of the above mentioned phenomena, as is seen in Figure 4.15, almost no organic substances are detected in the effluent which has passed through activated sludge, alum or lime coagulation with sedimentation and sand filtration, and the activated carbon adsorption process arranged in series.

Inorganic impurities in the Zone 4 are effectively removed by the processes of electrodialysis, ion exchange and so on. For the portion, electric conductivity is available to use as a comprehensive water quality index.

4.3 Conclusion

For the removal of organics from wastewater, three comprehensive water quality indices such as TOC, E260 mμ, and E220 mμ connected with impurity size distributions could describe the treatability of various types of treatment process such as biological process, chemical coagulation, carbon adsorption, and so on.

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3) Tambo,N., Kamei,T., and Uasa,A., Behaviors of organic pollutants in the physico-chemical treatment processes, Committee report on higher order water and wastewater controls, Japan society of civil engineering, Sanitary engineering committee (1971)

4) Standard method for the examination of water and wastewater, 13th edition, APHA, AWWA, WPCF (1971)

AGENDA

AMERICAN PRESENTATIONS

TUESDAY, FEBRUARY 12:

THE FEDERAL WATER POLLUTION CONTROL ACT AMENDMENTS OF 1972
FEDERAL VIEWPOINTS

THE FEDERAL WATER POLLUTION CONTROL ACT AMENDMENTS OF 1972
STATE VIEWPOINTS

EPA OVERALL RESEARCH PROGRAM AND WASTEWATER TREATMENT RESEARCH

WEDNESDAY, FEBRUARY 13:

TREATMENT AND DISPOSAL OF SLUDGE FROM MUNICIPAL WASTEWATER PLANTS IN THE UNITED STATES

EXPERIENCES WITH SLUDGE HANDLING IN TEXAS

THURSDAY, FEBRUARY 14:

PHYSICAL-CHEMICAL NITROGEN REMOVAL WASTEWATER TREATMENT

SLUDGES GENERATED IN PHOSPHATE REMOVAL PROCESSES

EPA EXPERIENCES IN OXYGEN-ACTIVATED SLUDGE

AERATION SYSTEMS FOR METRO CHICAGO

OXYGEN ACTIVATED SLUDGE SYSTEMS IN TEXAS

SUSPENDED SOLIDS REMOVAL PROCESSES STUDIED AT METRO CHICAGO

METRO CHICAGO - STUDIES ON NITRIFICATION

STORM AND COMBINED SEWER ABATEMENT TECHNOLOGY IN THE UNITED STATES
- AN OVERVIEW -

THE FEDERAL WATER POLLUTION CONTROL ACT AMENDMENTS OF 1972

FEDERAL VIEWPOINTS

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MUNICIPAL WASTE WATER SYSTEMS DIVISION
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PRESENTED AT

THIRD U.S./JAPAN CONFERENCE ON SEWAGE TREATMENT TECHNOLOGY
TOKYO, JAPAN

FEBRUARY 1974

THE FEDERAL WATER POLLUTION CONTROL ACT AMENDMENTS OF 1972

FEDERAL VIEWPOINTS

INTRODUCTION

In the United States, improving and maintaining the quality of our waters requires a total national commitment at all levels. For a long time in the United States water pollution was considered to be a matter for only local concern. Since 1956, however, the Federal government and the States have been actively assisting local communities with funds and technical assistance. Still this effort was not enough. The pollution of the waters grew to be a larger problem chiefly because of increased urbanization and industrialization.

Faced with this situation, our Congress developed the Federal Water Pollution Control Act Amendments of 1972 to provide planning and actions to deal with water pollution. This legislation covers a wide range of activities. However, the objectives and goals are expressed in the very first paragraphs of the Act, to wit:

1. By 1985, elimination of discharges of pollutants into navigable waters.
2. By 1983, the attainment of water quality which provides for the protection and propagation of fish, shellfish, and wildlife, and recreation in and on the waters.
3. Prohibition of the discharge of toxic pollutants in toxic amounts.

4. Federal financial assistance for the construction of publicly owned waste treatment works.
5. Fostering of areawide waste treatment management planning processes.
6. Maintaining a major research and demonstration effort to develop technology necessary to eliminate the discharge of pollutants into the navigable waters, waters of the contiguous zone, and the ocean.

At our last conference in December of 1972 we were just beginning to implement the Act. Since then we have worked very hard, learned much and in doing so have made significant progress toward full implementation of the Act. Today, I would like to describe to you our efforts over the past year. Since my specific area of responsibility is in municipal wastewater treatment technology, I will emphasize that subject area while giving a broad overview of the other portions of EPA's responsibility under the Act.

EFFLUENT LIMITATIONS

In general, the new legislation provides for definite effluent limitations for discharges into receiving waters, rather than application of the former provisions whereby water quality standards for the particular waters governed. Municipalities must have secondary treatment of their wastewaters by July 1977 or 1978 under the law. However, where wastewaters given secondary treatment will not achieve the water quality standards of the receiving body of water, higher than secondary treatment must be provided. This means, for example, that a State may set higher standards than secondary treatment for a particular lake, river, or estuary. These higher standards will determine the degree of treatment that must be provided to wastewaters discharged to the waterway.

On August 17, 1973 EPA published a final regulation which establishes the definition for the achievement of secondary treatment. It requires, in general, 85 per cent removal or 30 mg/l of BOD and suspended solids, whichever is more stringent. All municipalities must provide treatment of their wastewaters, at least to this standard, by July 1977 with some extensions to 1978 in the case of plants that are under construction.

The law also provides that no grant can be made with 1975 funds (available January 1, 1974) unless the treatment process involves the use of the best practicable technology currently available and further -- all municipalities, whether or not they receive grants, must have the best practicable treatment by 1983. We have drafted the best practicable technology regulation and it is now in the final stages of approval for publication as a proposal. We have provided you with copies and welcome your comments and suggestions. Basically, the regulation establishes criteria for three basic alternative waste management techniques which must be considered and evaluated for cost-effectiveness. The alternatives and criteria are summarized as follows:

Treatment and Discharge

	Monthly/Weekly Averages	
BOD_5	30/45 mg/l	
SS	30/45 mg/l	(Equivalent to Secondary Treatment)
Fecal Coliform	200/400/100 ml	
pH	(6-9)	

For treatment works over 1.0 MGD or 10,000 pop. or not discharing to the open ocean.

UOD	50/75 mg/l @ $20^{\circ}C^{+}$
UBOD	30/45 mg/l @ $20^{\circ}C$ -

$$UOD = 1.5(BOD_5) + 4.6(NH_3 \text{ as N}) - 1.0(D.O.)$$

$$UBOD = 1.5(BOD_5 - 1.0(D.O.)$$

Land Application

Permanent Ground Water Criteria

Chemical Parameters	EPA Drinking Water Supply Standards
Pesticides/Organics	EPA Drinking Water Supply Standards

Point Source Discharge Criteria

Same as Treatment and Discharge Levels

Reuse

Not to Exceed the Above Levels

In the case of industrial discharges, the effluent limitations to be imposed must be derived from effluent guidelines now being developed for 28 basic industrial categories that contribute significantly to water quality problems. These guidelines for industry put into definite terms the 1972 law's goals, i.e. "best practicable technology" by 1977 and then "best available technology" by 1983.

PERMITS

The Act establishes a new system of permits for discharges into the Nation's waters, replacing the 1899 refuse Act permit program. No discharge from any point source is allowed without a permit. These must be obtained for publicly owned treatment plants as well as industrial dischargers. Over 20,000 municipal treatment plants must obtain permits by December 1974. A typical permit will contain a schedule for upgrading treatment to come within prescribed effluent limitations.

Any discharge not in conformity with a permit will be unlawful, and, if willful and negligent, will be subject to a penalty of from $2500 to $25,000 per day of violation. On the part of EPA, court actions will be used as a last resort. We propose to give industries and municipalities the ample opportunity to comply with the requirements on a voluntary basis.

PRETREATMENT STANDARDS AND GUIDELINES

The pretreatment standards can be described in terms of the two major objectives outlined by Congress.

The first objective, to prevent the introduction of pollutants which would pass through inadequately treated, requires that the term "inadequately treated" be defined. The regulations are based on the premise that incompatible pollutants introduced by an industry are inadequately treated if they pass through publicly owned treatment works in amounts greater than would be permitted if the user discharged directly to the receiving waters. Accordingly, the pretreatment standards for incompatible pollutants are the same as the requirements for direct discharge. These requirements are to be based upon application of the best practicable control technology currently available.

An incompatible pollutant is defined as any pollutant other than BOD, suspended solids, pH and fecal coliform bacteria plus those pollutants that the municipal plant was specifically designed to treat. A less stringent pretreatment standard is permitted for thos incompatible pollutants which the municipality is committed to remove in its permit for the publicly owned treatment works.

The standards for incompatible pollutants apply only to major industries so as to reduce the number of industries which would have to pretreat and yet cover those industries that could have a significant impact upon the municipal plant.

The second objective which the standards must address is to protect the operation of the publicly owned treatment works. Under this objective, four prohibited discharges are listed in the standards.

There will be situations when the Federal pretreatment standards will not be sufficient to protect the operation of the publicly owned treatment works or prevent the discharge of industrial pollutants inadequately treated. In such cases, the municipality would have to supplement the Federal Standards. The pretreatment guidelines are intended to assist municipalities in accomplishing this.

CONSTRUCTION GRANT REGULATIONS

To enable municipalities to meet effluent limitations and permit requirements, the new amendments authorize Federal grants to municipalities for assistance in building sewage treatment facilities. These grants are mandated at the rate of 75 percent of eligible costs of approved projects.

To facilitate the processing of these grants, EPA has issued the "Title II Regulations." These regulations set forth policies and procedures concerning the processing of applications for grant assistance. Included is the provision that all projects must meet planning requirements and receive a priority certification from the State.

The regulations have been written with a view toward giving communities and States as much autonomy in making decisions as is possible under law. For example, the responsibility for review of plans and specifications for projects is to be passed to States as rapidly as States become able.

An innovation in the regulations is the introduction of a three-step grant process. Step 1 allows a separate grant for the preparation of preliminary studies and engineering. Step 2 provides for a grant for the preparation of construction drawings and specifications, and Step 3 is for a grant for the building and erection of the treatment works. This division of the financing of a grant for a project will accelerate payments to the communities and allow available funds to be spread over a larger number of projects. We believe that this procedure will be of great assistance to communities. It has already allowed a larger number of projects to go forward in most areas than would have been the case under the former regulations and law.

Final "Title II Regulations" will be published soon.

I would like to briefly mention some other requirements under the Act that you will be interested in:

PLANNING

The law lays down firm requirements for the planning of pollution abatement programs, and for control programs tied directly to the plans. For example, each State must have a continuing planning process which will result in water quality control plans for all navigable waters within the State. Included in such plans must be an inventory and ranking, in order of priority, of needs for construction of wastewater treatment plants required to meet the applicable standards. In addition, each sewage treatment works must have a facility plan which will consider not only the technical but also the social and economic aspects of a project. Involved here are environmental impact statements prepared by EPA when major controversies are unresolved.

INFILTRATION/INFLOW

Section 201 of the Act states that the EPA Administrator shall not approve a grant after July 1, 1973 unless the applicant shows that each sewer system discharging into the treatment works is not subject to excessive infiltration. The construction grant regulations require that an analysis be made of the sewer system involved to determine if there are indications of excessive infiltration. If so, an infiltration study has to be made to determine action to handle the infiltration problem economically. Such studies are eligible grant costs, usually as part of the Step 2 grant process. There will be cases, where it will be more economic, on a cost-effective basis to treat infiltration by building a larger plant, rather than trying to seal a sewer system. Also, it is recognized that in many cases it may not be feasible to remedy infiltration/inflow immediately. A reasonable and satisfactory abatement schedule may be agreed upon while the project goes forward.

An infiltration/inflow analysis has always been part of a well-planned project. Therefore, we do not look upon this requirement as being burdensome in the design of sewage treatment systems. Guidelines on infiltration/inflow analysis will be published shortly at about the same time as final construction grant regulations.

COST-EFFECTIVENESS

Section 212 of the Act also specifies that the Administrator shall publish guidelines on methods of cost-effectiveness analysis for the construction of treatment works. These guidelines were published in final form on September 10, 1973. Basically, the guidelines contain uniform economic analysis procedures which must be incorporated into all grant applications. This will assure adequate data and analysis demonstrating the project to be the most cost-effective over the design life of the works. The analysis must include consideration of both capital and operation and maintenance costs.

Again, cost effectiveness analysis has normally been undertaken as part of any sound engineering analysis, even though it may not have been labeled as such. The law formalizes the procedures. The requirement is not expected to cause any great problems since the process is familiar to those in the profession.

USER CHARGES AND INDUSTRIAL COST RECOVERY

Having provided for Federal assistance in building wastewater treatment facilities, the Act seeks to assure that the facilities will be properly maintained and operated. To accomplish this, communities that are assisted with grants are required to have a user charge system that insures that all users will pay their proportionate share of operation and maintenance costs. In addition, the law provides that industry, discharging into the system, must pay back its proportionate share of the Federal grant. Fifty percent of this payback may be used for liquidating the community's cost of the project and for future expansion and reconstruction of the project. The remaining 50 percent reverts to the U.S. Treasury. Final user charge and cost recovery regulations were published on August 21, 1973.

FINANCIAL NEEDS OF THE PROGRAM

To achieve the goals of the program will be a costly venture. Recent estimates indicate that it will require industry to expend for capital improvements $12.9 billion to bring about the use of best practicable technology by 1977 and an additional $7.8 billion to apply best economically available technology by 1983.

The needs of municipalities for financing has been the subject of a recent survey. The objective of the survey was to ascertain the funds that would be required to meet the 1977 goals of attaining secondary treatment or higher where water quality standards are higher. The total amount came to a total of over $60 billion. $16.6 billion is required to improve treatment plants to achieve secondary treatment, with an additional $5.6 billion for further removing specific pollutants such as phosphorus, nitrogen and organics to the extent required by legally binding Federal, State, international and local actions. $700 million is required for eliminating infiltration/inflow conditions, $13.6 billion is necessary for building new interceptors, force mains, and pumping stations, $11 billion is estimated for new collection systems, and $12.7 billion is required for the reduction of combined sewer overflows. We know that some of these figures are not absolutely accurate because of an inadequate data base that we are even now engaged in strengthening. We believe, however, that the combined total of $36 billion for treatment plants, removal of other pollutants, and the building of new interceptors is reasonably accurate. The remaining $24 billion is probably too low a figure.

FEDERAL FINANCIAL ASSISTANCE

The Federal government between 1956 and 1972 expended over $5 billion to assist municipalities in building waste treatment facilities. At present another $9 billion is available for grants at 75 percent of construction costs. Four billion of this total was just added on January 1, 1974. Further financing for after that date is now under study and review.

CONCLUSION

The magnitude of the United States' program which I have just described leads to my concluding remarks. To make such a program viable will require the utmost use of new technology and innovations. Even samll advances in improving the efficiency and effectiveness of sewage treatment systems will result in major savings overall; significant advances will reduce the load on our taxpayers who must pay for the improvements in water quality.

Your efforts in Japan to improve and maintain water quality must present the same strains on your economic and social systems. By joining together to bring about the most efficient and effective application of technology and ideas to the treatment of wastewaters, we can achieve more than each of us can alone. Our visit to Japan so far has been most beneficial in this respect. I look forward to a successful conclusion of this Conference.

THE FEDERAL WATER POLLUTION CONTROL ACT AMENDMENTS OF 1972

STATE VIEWPOINT

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TOKYO, JAPAN

FEBRUARY 1974

FEDERAL WATER POLLUTION CONTROL ACT AMENDMENTS OF 1972--STATE VIEWPOINT

I believe it appropriate in the beginning to provide a limited overview of the American system of state-federal relations inasmuch as this will help you understand my view and the interactions of federal, state, and local governments in the United States.

The U.S. Constitution, written by representatives from the several states, provides for a division of governmental powers between the federal and state governments. The division was accomplished by delegating to the federal government specified powers plus the powers inherently required to execute the delegated powers--all undelegated powers to remain with the states.

Under this division of powers, the control of water pollution was for many years considered to be primarily a function of state government. The federal government was considered as having the prime function when its ability to discharge its delegated powers were endangered. Along these lines, the federal government in 1899 enacted a refuse act which placed it into the water pollution effort in a very limited way--only to protect navigation channels from excessive siltation or clogging. The 1899 Refuse Act required a federal permit to discharge refuse-laden wastewaters into navigable waters. Several years ago, as a result of a judicial decision, the 1899 Refuse Act was expanded in scope, wrongly in the opinion of many, to consider all pollutants--not only suspended solids. The 1972 Amendments, at least to some degree, is an outgrowth of the 1899 Refuse Act since it sets up a comprehensive federal waste discharge permit system.

The U.S. Constitution has generally been interpreted so as to provide federal supremacy over state government when there is a conflict involving the national interest. This is the case with the 1972 Amendments where it was decided that the national interest was being endangered because of the failure of the several states as a whole to adequately control water pollution. When this is done, a state's power to cope with its own problems as it sees fit is to a considerable degree set aside--even though a particular state may be doing its job satisfactorily.

Under the division of powers previously discussed, the State of Texas proceeded many years ago to inaugurate programs to control water pollution--other states also inaugurated programs, some very good, some very bad. I think Texas has a good program. We have required secondary treatment of domestic sewage for some 30 years. We have had a permit program covering both municipal and industrial discharges, including agriculture, since 1962. We have had a mandatory self-monitoring program in operation since 1970. All of these programs and more are now incorporated into the 1972 Amendments and made federal programs. This fact, coupled with detailed regulations promulgated under the Act requires that all our programs be revamped and restructured to fit the pattern mandated by the federal government. This necessarily involves a great deal of lost motion and unnecessary expense. In this case, states who did not have comprehensive programs as of the passage of the Act are probably in an advantageous position since they will be able to structure programs from the beginning to fit the federal mold. As each of you know, it is easier to construct a new house to detailed specifications than it is to rebuild an old one--even though the old structure may be sound and sturdy. We are in that position in Texas and we are trying to rebuild our house since we recognize the Act as the law of the land.

I think it would be appropriate at this time to say that the various state governments have no single viewpoint on the merits of the 1972 Act. This same divergence of opinion exists even within individual states between state civil servants, the political leadership, and the people. In general, I think it would be true that: (1) the state civil servants, almost to a man, think the law has serious flaws and has been wrongly implemented; (2) the political leadership recognizes the law as a political necessity; and (3) the people have no specific view--they are merely dissatisfied with the progress the states have made in correcting pollution problems. In these regards, I shall only attempt to discuss the Texas viewpoint, generally as seen by state pollution control officials.

Within the time allotted to me, let us look at some aspects of the law or regulations promulgated therunder which we consider problems.

Section 101(b) states that it is the intent of Congress that the rights of the states to protect and use its waters be preserved. In actuality, the federal role is all pervasive and conformity to the federal mold is much more mandatory than statutory language would require. This, we think, is a mistake since it fails to take into account regional differences between the states in law, custom, history, land use, topography, etc. The insistence on uniformity has caused considerable wear and tear on nerves and tempers, and has served no good purpose.

Section 101(a) sets forth the goals of the Act--recreational use water quality in all streams by 1983 and no discharge of pollutants by 1985. I think these goals are unrealistic, particularly the schedules. Such environmental daydreaming will never take the place of common sense and sound technical judgment.

Section 301(b) (1) (c) of the Act says in effect that by July 1, 1977, publicly owned treatment works shall produce either a secondary treated effluent, or an effluent subject to advanced waste treatment if this is required to meet stream standards. Federal waste discharge permits as provided for in Section 402 of the Act will be used to implement these requirements including the schedules. So far, all is well and proper, except possibly the schedule.

When, however, the permit program is coupled with the grant program for publicly owned treatment works as established under Title II of the Act--a real dilemma results.

Section 202(a) provides that the federal shore of the costs of publicly owned treatment works shall be 75%. A survey conducted under Section 516(b) (2) of the Act to determine the needs of all the states resulted in a need for Texas in excess of 800 million dollars, and we think at the state level this is too low. The actual level of funding as evidenced by past actions will fall far short of the need. Thus, taking into consideration both the permit program and the grant program, it would appear to us that the federal government is telling state and local government on the one hand through the permit system to construct rapidly the needed works; and on the other hand telling these same governments through the grant program that they are entitled to 75% federal funding, but that it will not be forthcoming in many cases in time to meet the 1977 deadline. This in our judgment is confusing. We are very hopeful that the Congress will resolve this problem so that our nation can proceed harmoniously to construct the necessary works to finish the task of cleaning up our waters.

Title IV of the Act provides for a permit program to control wastewater discharges, and provides that the federal government can delegate under certain conditions the responsibility for the administration of this program to the states. EPA has promulgated restrictive regulations concerning the prerequisite to receive delegation and has interpreted the regulations narrowly. They appear to be attempting to find reasons why a state should not be given delegation, rather than how to overcome obstacles to delegation. We think this is a mistake and that this fails to properly utilize existing trained manpower available in state organizations. In order to partially overcome this problem, our state and the EPA Region in which we are located have worked out an informal arrangement to make our resources available.

Many other issues could be raised; however, suffice it to say that we think that inadequate attention has been given in the law and its implementation toward effectively utilizing existing state organizations. This is, we believe, primarily due to an attempt to fit all state programs--good and bad--into one rigid mold.

Contrary to what you may think at this point in my presentation, I do not think all aspects of the Act are bad. The concept of setting water quality standards for the nation's waters is sound and necessary.

The concept of specifying minimum treatment levels regardless of stream requirements is sound--we have followed this concept in Texas in the case of municipal sewage for over 30 years. The continuing planning process mandated by Section 303(e) of the Act is a sound concept. It has enabled us to take a more effective look at the overall water quality needs in each river basin and will result in a much more coordinated control program than has heretofore been the case. The self-monitoring program wherein each waste discharger is required to monitor the quantity of pollutants he discharges to public waters and report this information to the government is sound. We have had such a program in operation in Texas since 1970 and it has been one of the most effective tools we have found in recent years in abating pollution. There are many other aspects of the Act which are worthwhile and necessary.

In summary, let me say that while I have obvious concerns about the Act itself and the manner in which EPA has implemented the Act, it is a far-reaching Act clearly setting forth our government and our people's determination to solve our nation's water pollution problems. I am personally very confident that in a short time the rough spots will be smoothed over, and we will complete the job of cleaning our nation's waters, started so many years ago. I started in this business over 20 years ago, and I am glad to see the climate of public opinion change to support the clean-up effort. Without the public's support, the job will not be done. The passage of the 1972 Act is very significant in that it clearly signals our nation's commitment to clean water.

EPA OVERALL RESEARCH PROGRAM
AND WASTEWATER TREATMENT RESEARCH

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TOKYO, JAPAN

FEBRUARY 1974

EPA OVERALL RESEARCH PROGRAM AND WASTEWATER TREATMENT RESEARCH

ORGANIZATION

The Environmental Protection Agency is headed by an Administrator, now Mr. Russell E. Train. To carry out the EPA programs there are Assistant Administrators for various programs as shown in Chart I.

Research is directed by the Assistant Administrator for Research and Development, Dr. Stanley M. Greenfield. His office is in Washington, D.C. The Office of Research and Development has about 1800 employees and a budget of about $120,000,000 per year. Chart II shows how the Office is organized.

To conduct the research for EPA, four National Environmental Research Centers have been established. These NERC's are located at Cincinnati, Ohio; Corvallis, Oregon; Research Triangle Park, North Carolina and Las Vegas, Nevada. There are about ten smaller laboratories in other locations that are administered by the NERC's. The brochures you have been provided explain the programs of the NERC's further. Chart III shows the organization of the Cincinnati NERC - The Robert A. Taft Laboratory.

HOW OUR RESEARCH IS CONDUCTED

Wastewater treatment research is conducted by the Advanced Waste Treatment Research Laboratory in Cincinnati. Industrial waste treatment research is conducted at Edison, New Jersey and at some other NERC's. Our Storm and Combined Sewer Research Program is also located at Edison, New Jersey.

About 150 people work in the programs devoted to domestic and industrial waste treatment. The budget for domestic wastewater is about $10,000,000 and the industrial waste budget is about the same.

The objectives of our treatment research programs are to improve old methods and devise new methods for the management and control of wastewaters to enhance the quality of the Nation's water and meet the quality standards required.

How do we decide upon research projects? First of all, the EPA is guided by certain laws that have been enacted. Much of our work in water pollution is directed by the Federal Water Pollution Amendments Act of 1972 - Public Law 92-500. Within the policies of the EPA, the Office of Research and Development develops a basic strategy to meet the Agency needs. Research needs are solicited from EPA Headquarters and field programs. Using these needs and the strategy documents, our Washington Headquarters, with help from the laboratories, draws up a general research plan and designates the amount of money to be spent in each area of work. It is then the job of the research staff to design detailed work plans. After the work plans are approved, the project gets underway. We project our research for five to seven years. We perform research in our own laboratories but we also make extensive use of cities, industries, commercial research groups and universities to conduct research for us. Chart IV shws our planning process in graphical form. Chart V is a list of major topics we are conducting research on in the municipal and industrial waste field. Chart VI shwos a summary sheet on one project and a graphical network diagram of the project.

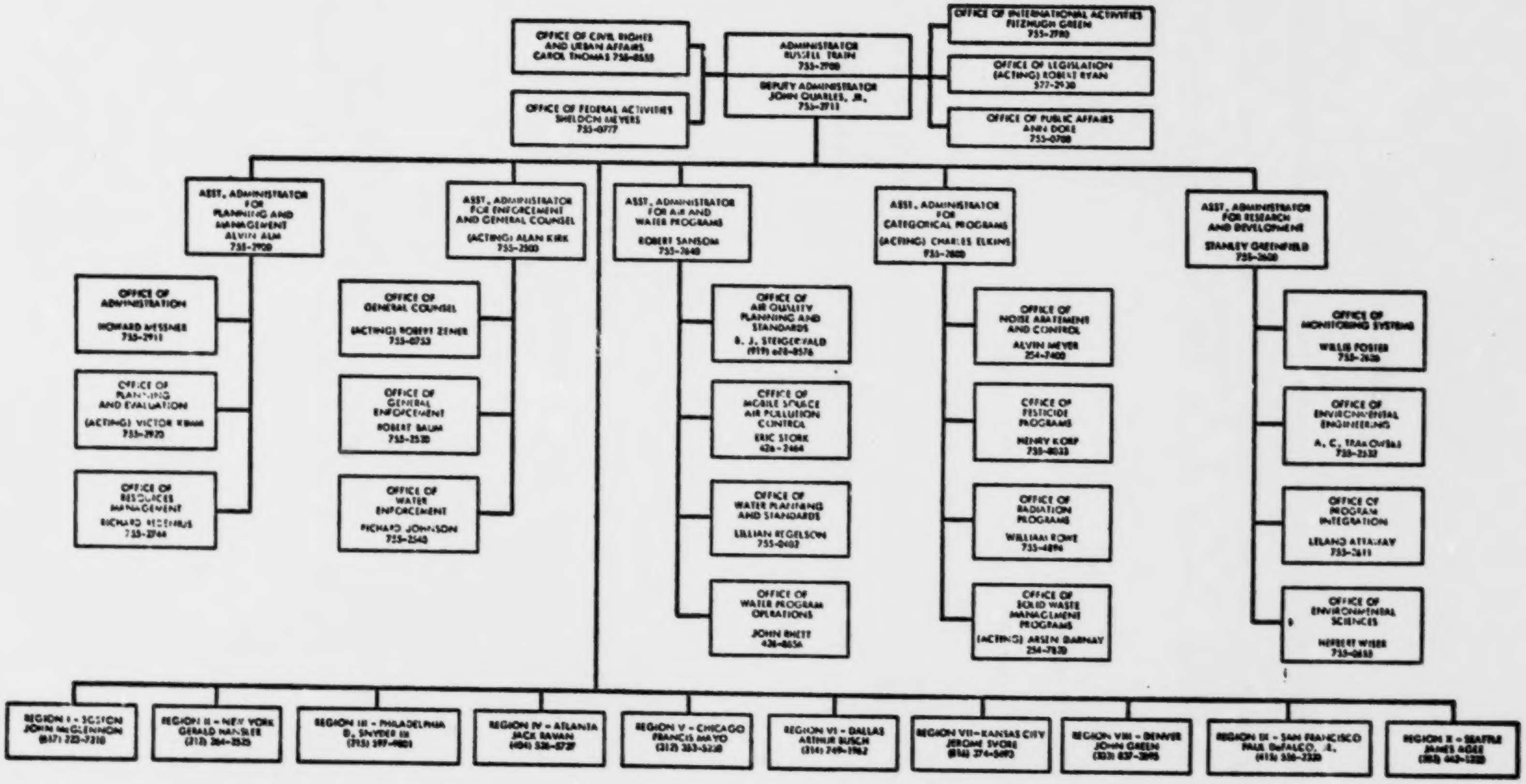
TECHNOLOGY TRANSFER

Once the research is completed, it is made available to the Agency programs that has requested it and to others. There are now several hundred reports on the research that has been done. These reports are available to you.

To expedite the use of new technology, EPA has special teams of researchers and outside consultants put on special two- or three-day seminars all over the country for consulting engineers, state authorities, city governments and the like. We also produce manuals on various subjects and distribute them widely. You are already familiar with our Advanced Waste Treatment Manuals.

CHART I

U. S. ENVIRONMENTAL PROTECTION AGENCY


ADMINISTRATOR RUSSELL TRAIN 755-2700
DEPUTY ADMINISTRATOR JOHN QUARLES, JR. 755-2711
OFFICE OF CIVIL RIGHTS AND URBAN AFFAIRS CAROL THOMAS 755-0555
OFFICE OF FEDERAL ACTIVITIES SHELDON MEYERS 755-0777
OFFICE OF INTERNATIONAL ACTIVITIES FITZHUGH GREEN 755-2780
OFFICE OF LEGISLATION (ACTING) ROBERT RYAN 577-2930
OFFICE OF PUBLIC AFFAIRS ANN DORE 755-0700
ASST. ADMINISTRATOR FOR PLANNING AND MANAGEMENT ALVIN ALM 755-2900
OFFICE OF ADMINISTRATION HOWARD MESSNER 755-2911
OFFICE OF PLANNING AND EVALUATION (ACTING) VICTOR KIMM 755-2920
OFFICE OF RESOURCES MANAGEMENT RICHARD REDENIUS 755-2744
ASST. ADMINISTRATOR FOR ENFORCEMENT AND GENERAL COUNSEL (ACTING) ALAN KIRK 755-2500
OFFICE OF GENERAL COUNSEL (ACTING) ROBERT ZENER 755-0753
OFFICE OF GENERAL ENFORCEMENT ROBERT BAUM 755-2530
OFFICE OF WATER ENFORCEMENT RICHARD JOHNSON 755-2540
ASST. ADMINISTRATOR FOR AIR AND WATER PROGRAMS ROBERT SANSOM 755-2640
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OFFICE OF MOBILE SOURCE AIR POLLUTION CONTROL ERIC STORK 426-2464
OFFICE OF WATER PLANNING AND STANDARDS LILLIAN REGELSON 755-0402
OFFICE OF WATER PROGRAM OPERATIONS JOHN RHETT 426-8856
ASST. ADMINISTRATOR FOR CATEGORICAL PROGRAMS (ACTING) CHARLES ELKINS 755-2800
OFFICE OF NOISE ABATEMENT AND CONTROL ALVIN MEYER 254-7400
OFFICE OF PESTICIDE PROGRAMS HENRY KORP 755-8033
OFFICE OF RADIATION PROGRAMS WILLIAM ROWE 755-4894
OFFICE OF SOLID WASTE MANAGEMENT PROGRAMS (ACTING) ARSEN DARNAY 254-7820
ASST. ADMINISTRATOR FOR RESEARCH AND DEVELOPMENT STANLEY GREENFIELD 755-2600
OFFICE OF MONITORING SYSTEMS WILLIS FOSTER 755-2626
OFFICE OF ENVIRONMENTAL ENGINEERING A. C. TRAKOWSKI 755-2532
OFFICE OF PROGRAM INTEGRATION LELAND ATTAWAY 755-2611
OFFICE OF ENVIRONMENTAL SCIENCES HERBERT WISER 755-0650
REGION I - BOSTON JOHN McGLENNON (617) 223-7210
REGION II - NEW YORK GERALD HANSLER (212) 264-2525
REGION III - PHILADELPHIA D. SNYDER III (215) 597-9801
REGION IV - ATLANTA JACK RAVAN (404) 526-5727
REGION V - CHICAGO FRANCIS MAYO (312) 353-5250
REGION VI - DALLAS ARTHUR BUSCH (214) 749-1962
REGION VII - KANSAS CITY JEROME SVORE (816) 374-5493
REGION VIII - DENVER JOHN GREEN (303) 837-3895
REGION IX - SAN FRANCISCO PAUL DeFALCO, JR. (415) 556-2320
REGION X - SEATTLE JAMES AGEE (206) 442-1220

CHART II

OFFICE OF RESEARCH AND DEVELOPMENT

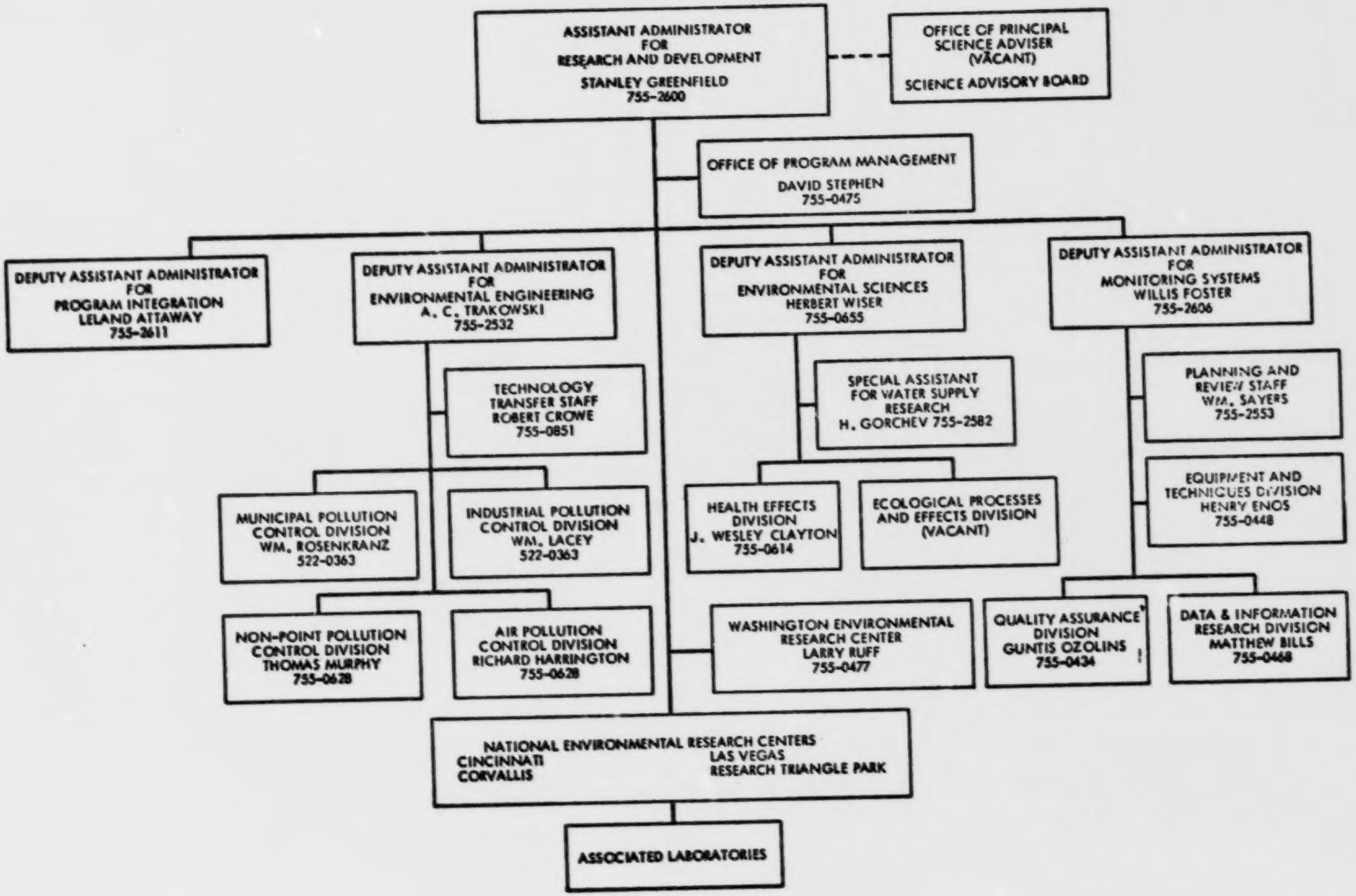
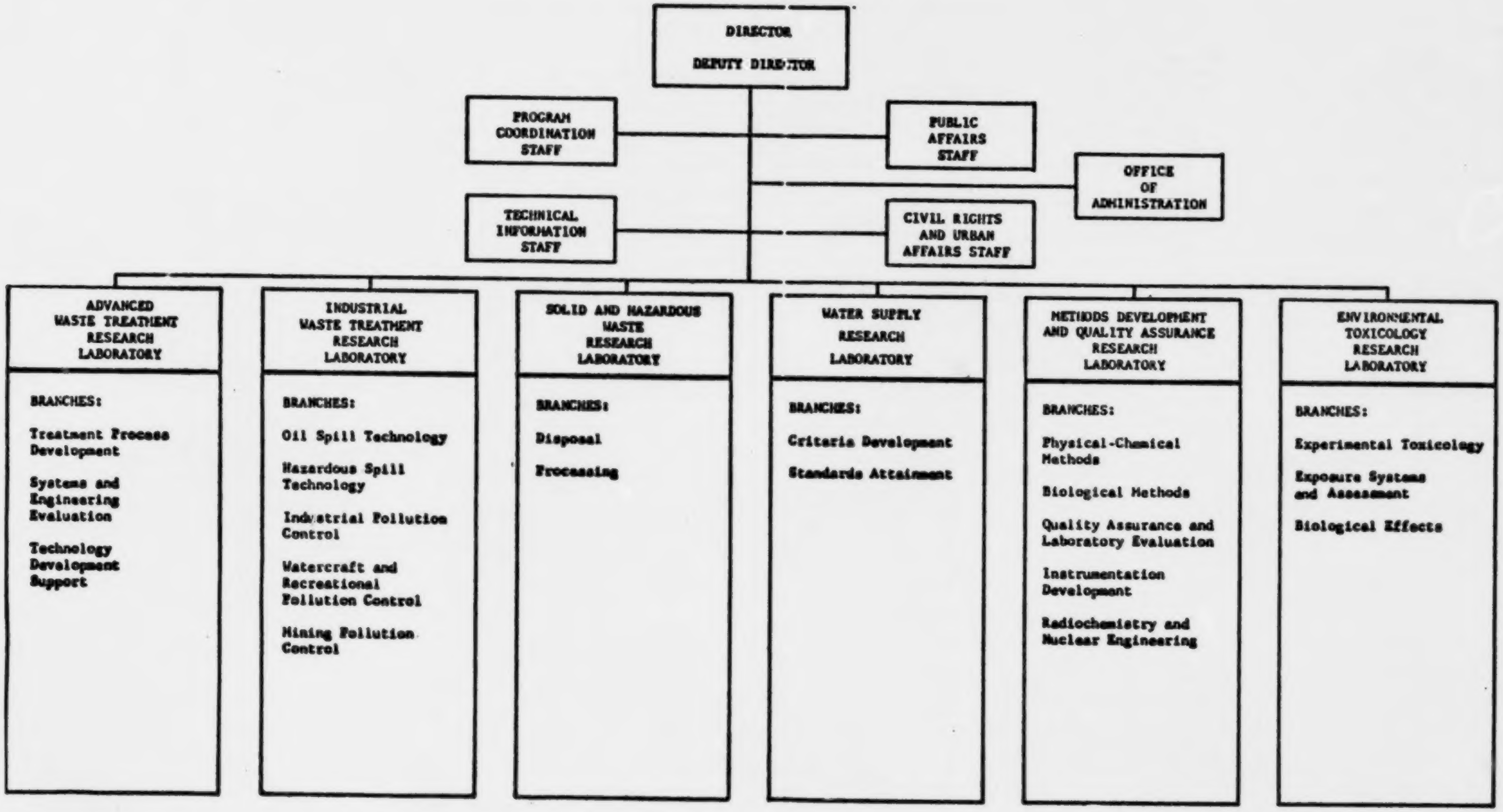

ASSISTANT ADMINISTRATOR FOR RESEARCH AND DEVELOPMENT STANLEY GREENFIELD 755-2600
OFFICE OF PRINCIPAL SCIENCE ADVISER (VACANT) SCIENCE ADVISORY BOARD
OFFICE OF PROGRAM MANAGEMENT DAVID STEPHEN 755-0475
DEPUTY ASSISTANT ADMINISTRATOR FOR PROGRAM INTEGRATION LELAND ATTAWAY 755-2611
DEPUTY ASSISTANT ADMINISTRATOR FOR ENVIRONMENTAL ENGINEERING A. C. TRAKOWSKI 755-2532
DEPUTY ASSISTANT ADMINISTRATOR FOR ENVIRONMENTAL SCIENCES HERBERT WISER 755-0655
DEPUTY ASSISTANT ADMINISTRATOR FOR MONITORING SYSTEMS WILLIS FOSTER 755-2606
TECHNOLOGY TRANSFER STAFF ROBERT CROWE 755-0851
SPECIAL ASSISTANT FOR WATER SUPPLY RESEARCH H. GORCHEV 755-2582
PLANNING AND REVIEW STAFF WM. SAYERS 755-2553
EQUIPMENT AND TECHNIQUES DIVISION HENRY ENOS 755-0448
MUNICIPAL POLLUTION CONTROL DIVISION WM. ROSENKRANZ 522-0363
INDUSTRIAL POLLUTION CONTROL DIVISION WM. LACEY 522-0363
HEALTH EFFECTS DIVISION J. WESLEY CLAYTON 755-0614
ECOLOGICAL PROCESSES AND EFFECTS DIVISION (VACANT)
NON-POINT POLLUTION CONTROL DIVISION THOMAS MURPHY 755-0628
AIR POLLUTION CONTROL DIVISION RICHARD HARRINGTON 755-0628
WASHINGTON ENVIRONMENTAL RESEARCH CENTER LARRY RUFF 755-0477
QUALITY ASSURANCE DIVISION GUNTIS OZOLINS 755-0434
DATA & INFORMATION RESEARCH DIVISION MATTHEW BILLS 755-0468
NATIONAL ENVIRONMENTAL RESEARCH CENTERS CINCINNATI CORVALLIS LAS VEGAS RESEARCH TRIANGLE PARK
ASSOCIATED LABORATORIES

CHART III

ORGANIZATION FOR NATIONAL ENVIRONMENTAL RESEARCH CENTER
THE ENVIRONMENTAL PROTECTION AGENCY
CINCINNATI, OHIO


DIRECTOR
DEPUTY DIRECTOR
PROGRAM COORDINATION STAFF
PUBLIC AFFAIRS STAFF
OFFICE OF ADMINISTRATION
TECHNICAL INFORMATION STAFF
CIVIL RIGHTS AND URBAN AFFAIRS STAFF
ADVANCED WASTE TREATMENT RESEARCH LABORATORY
BRANCHES:
Treatment Process Development
Systems and Engineering Evaluation
Technology Development Support
INDUSTRIAL WASTE TREATMENT RESEARCH LABORATORY
BRANCHES:
Oil Spill Technology
Hazardous Spill Technology
Industrial Pollution Control
Watercraft and Recreational Pollution Control
Mining Pollution Control
SOLID AND HAZARDOUS WASTE RESEARCH LABORATORY
BRANCHES:
Disposal
Processing
WATER SUPPLY RESEARCH LABORATORY
BRANCHES:
Criteria Development
Standards Attainment
METHODS DEVELOPMENT AND QUALITY ASSURANCE RESEARCH LABORATORY
BRANCHES:
Physical-Chemical Methods
Biological Methods
Quality Assurance and Laboratory Evaluation
Instrumentation Development
Radiochemistry and Nuclear Engineering
ENVIRONMENTAL TOXICOLOGY RESEARCH LABORATORY
BRANCHES:
Experimental Toxicology
Exposure Systems and Assessment
Biological Effects

September 14, 1973

CHART IV

OFFICE OF RESEARCH AND DEVELOPMENT PLANNING PROJECTS

EPA POLICIES AND
LEGAL RESPONSIBILITIES

↓

OFFICE OF RESEARCH AND DEVELOPMENT
RESEARCH NEEDS RECEIVED FROM OTHER
EPA PROGRAMS, STATES, ETC.

↓

STRATEGY PLAN
CONCURRED IN BY OTHER EPA PROGRAMS
GENERAL PROJECT PLANS
MONEY ESTIMATES PREPARED

↓

LABORATORY RESEARCH PERSONNEL
DESIGN DETAILED STUDIES
APPROVAL OF PLANS

WORK BEGINS

CHART V

RESEARCH PROJECTS IN DOMESTIC AND INDUSTRIAL WASTE TREATMENT AND CONTROL
NERC-CINCINNATI
1974

Advanced Waste Treatment Research Laboratory

Title
Demonstrate Combinations of Processes to Meet Water Quality Needs
Methods & Processes to Provide Improved Operation & Maintanence, In-System treatment, & Treatment of Joint Municipal & Industrial Wastes
Research & Investigation of Joint Liquid-Solid Waste Collection & Treatment
Treatment of Combined Sewer Overflows & Storm Water Discharges
Technology for Hydraulic & Pollutant Control of Urban Runoff
Simulation Models for Total Management of Severage Systems
Methods, Processes, & Systems to Reduce Water Use & Total Sewage Flow
Research, Development, & Pilot Projects to Eliminate Pollution from Sewage in Rural or Other Areas Generating Small Flows
Wastewater Renovation and Reuse
Wastewater System Instrumentation & Automation
Wastewater Sludge Processing & Treatment
Wastewater Treatment Sludge Disposal
Control of Nutrients in Wastewater
Suspended & Colloidal Solids Removal from Municipal Wastewater
Biological Treatment Process Improvements for Municipal Wastewater Applications
Municipal Wastewater Disinfection Process Development and Demonstration
Reduction of Total Dissolved Solids (TDS) & Heavy Metals in Municipal Wastewaters
Control of Dissolved Organics in Municipal Wastewater by Physical-Chemical Processes

CHART V (continued)

Industrial Waste Treatment Research Laboratory

Technology Research for the Elimination of the Discharge of Pollutants from the Inorganic & Miscellaneous Chemicals Industries

Technology Research for the Elimination of the Discharge of Pollutants from the Non-Ferrous Metals & Electroplating Industries

Technology Research for the Elimination of the Discharge of Pollutants from the Rubber & Plastics Industries

Completion of FY-73 Work Plan ROAPs 21 APK, 21 APL and 21 APO

Treatment of Mine Drainage

Pollution Control Methods for Surface Mining

Control of Pollution from Underground Mining of Solid Fuels

New Mining Methods

Mine Water Pollution Control Demonstrations - Section 107

OHMSETT Support

Chemical Identification of Oil Spills

Oil Spill Containment Devices

Equipment for Physically Removing Oil Spilled in the Environment

Waste Oil Recycling

Prevention of Hazardous Material Spills

Hazardous Material Spill Emergency Response

Hazardous Material Spill Control and Removal

Separation & Recovery of Removed Spilled Hazardous Materials

Environmental Evaluation of Devices & Techniques to Control Hazardous Material Spills

Office of Research and Monitoring
RESEARCH OBJECTIVE ACHIEVEMENT PLAN SUMMARY

PROGRAM ELEMENT TITLE Treatment Process Development and Optimization	PROGRAM AREA TITLE Municipal Technology	ROAP/TASK NO. 21-AST PROGRAM ELEMENT NO. 1B2043
ROAP/TASK TITLE: Reduction of Total Dissolved Solids (TDS) and Heavy Metals in Municipal Wastewaters		FORMER ROAP NO(S). 21ABO 21AAQ 21AAP 21AAS

ROAP OUTPUT:

Design guidelines for cost-optimized reverse osmosis, ion exchange and distillation processes based on large-scale demonstrations. Reports on capabilities of waste treatment processes to remove heavy metals and possible process modifications to enhance removal of metals and toxic non-metals.

ROAP COMPLETION DATE: 1981

ROAP APPROACH:

Reverse osmosis, ion exchange, and distillation of wastewater will be demonstrated at 100,000 gpd scale. Concurrent projects will explore solutions to problems of brine disposal and regenerant reclamation. Capabilities of physical-chemical processes to remove toxic metals will be determined on 4 gpm pilot plants. Possible process modifications to enhance removal of metals will be examined.

PROGRESS TO DATE AND PROBLEMS ENCOUNTERED:

Reverse osmosis, ion exchange, and distillation have all been investigated at bench and small pilot scale. Costs are and will remain high, but some reduction should be possible through process optimization. Metal removal capabilities for 11 metals has been determined. Reconstruction of pilot plants has delayed continuation for additional metals.

MAJOR MILESTONE EVENTS (Prior & Planned)	TARGET DATES Planned	Actual
1. Start Date		1963
2. Complete first tests of reverse osmosis applied to wastewater	1964	1964
3. Laboratory evaluation of distillation of wastewater	1964	1964
4. Demonstrate feasibility of weak-base resin for wastewater treatment	1967	1967
5. Investigate Infilco 4-bed ion exchange process	1968	1968
6. Determine techn. feasibility of existing commercial reverse osmosis equipment for treating secondary effluent (continued)	1968	1968

ORGANIZATION RESPONSIBLE FOR PREPARING AND IMPLEMENTING ROAP

☐ ORM Hq. ☒ NERC, Cinc.
☐ NERC, Corv. ☐ NERC, RTP
☐ NERC, Las V.

LAB/DIV ORGANIZATION CODE (five digits): 67024

R&D RELATED PROGRAM ELEMENTS
Supporting PE No's:
Dependent PE No's:

TOTAL RESOURCES	FY's Previous	FY 73 Current	FY 74 Budget	FY 75	FY 76	FY 77	FY 78	FY 79	Total Cost of ROAP
Funding $(K)	3106	356	68	220	375	600	510	663	5898
Pers. M-Yrs.	40.7	1.2	2.3	2:3	2.3	3.7	2.3	0.4	55.5

NEED SPONSORS

Sponsor	Priority	Need ID
Region V	131/171	05 ABO
OAWP	17/62	16 AKU
NERC Cinti	191/454	24 AXP

PROGRAM AREA PRIORITY

Assigned By	Priority	Date

ROAP REVISIONS

Revision	Date of Revision
Prep Date	2/21/73
A	3/13/73
B	4/24/73
C	
D	
E	
F	

ORM Form R-1

CHART VI

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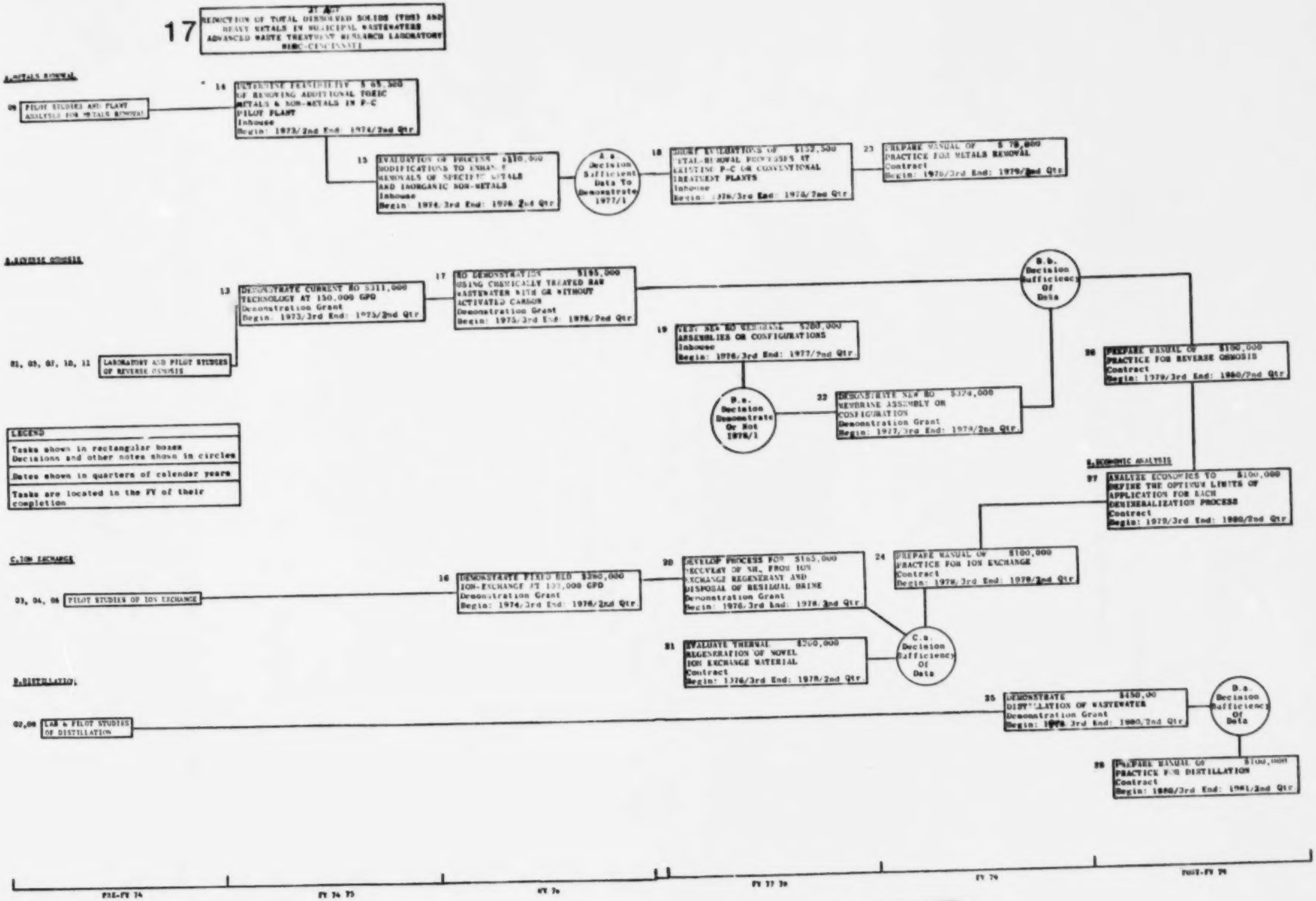
17
REDUCTION OF TOTAL DISSOLVED SOLIDS (TDS) AND
HEAVY METALS IN MUNICIPAL WASTEWATERS
ADVANCED WASTE TREATMENT RESEARCH LABORATORY
NERC-CINCINNATI
A. METALS REMOVAL
09 PILOT STUDIES AND PLANT ANALYSIS FOR METALS REMOVAL
14 DETERMINE FEASIBILITY $ 65,500 OF REMOVING ADDITIONAL TOXIC METALS & NON-METALS IN P-C PILOT PLANT Inhouse Begin: 1973/2nd End: 1974/2nd Qtr
15 EVALUATION OF PROCESS $110,000 MODIFICATIONS TO ENHANCE REMOVALS OF SPECIFIC METALS AND INORGANIC NON-METALS Inhouse Begin: 1974/3rd End: 1976/2nd Qtr
A.a Decision Sufficient Data To Demonstrate 1977/1
18 SHORT EVALUATIONS OF $132,500 METAL-REMOVAL PROCESSES AT EXISTING P-C OR CONVENTIONAL TREATMENT PLANTS Inhouse Begin: 1976/3rd End: 1978/2nd Qtr
23 PREPARE MANUAL OF $ 78,000 PRACTICE FOR METALS REMOVAL Contract Begin: 1978/3rd End: 1979/2nd Qtr
B. REVERSE OSMOSIS
01, 05, 07, 10, 11 LABORATORY AND PILOT STUDIES OF REVERSE OSMOSIS
13 DEMONSTRATE CURRENT RO $311,000 TECHNOLOGY AT 150,000 GPD Demonstration Grant Begin: 1973/3rd End: 1975/2nd Qtr
17 RO DEMONSTRATION $195,000 USING CHEMICALLY TREATED RAW WASTEWATER WITH OR WITHOUT ACTIVATED CARBON Demonstration Grant Begin: 1975/3rd End: 1976/2nd Qtr
19 TEST NEW RO MEMBRANE $200,000 ASSEMBLIES OR CONFIGURATIONS Inhouse Begin: 1976/3rd End: 1977/2nd Qtr
B.a. Decision Demonstrate Or Not 1978/1
22 DEMONSTRATE NEW RO $374,000 MEMBRANE ASSEMBLY OR CONFIGURATION Demonstration Grant Begin: 1977/3rd End: 1979/2nd Qtr
B.b. Decision Sufficiency Of Data
26 PREPARE MANUAL OF $100,000 PRACTICE FOR REVERSE OSMOSIS Contract Begin: 1979/3rd End: 1980/2nd Qtr
LEGEND
Tasks shown in rectangular boxes
Decisions and other notes shown in circles
Dates shown in quarters of calendar years
Tasks are located in the FY of their completion
E. ECONOMIC ANALYSIS
27 ANALYZE ECONOMICS TO $100,000 DEFINE THE OPTIMUM LIMITS OF APPLICATION FOR EACH DEMINERALIZATION PROCESS Contract Begin: 1979/3rd End: 1980/2nd Qtr
C. ION EXCHANGE
03, 04, 06 PILOT STUDIES OF ION EXCHANGE
16 DEMONSTRATE FIXED BED $280,000 ION-EXCHANGE AT 100,000 GPD Demonstration Grant Begin: 1974/3rd End: 1976/2nd Qtr
20 DEVELOP PROCESS FOR $165,000 RECOVERY OF NH_3 FROM ION EXCHANGE REGENERANT AND DISPOSAL OF RESIDUAL BRINE Demonstration Grant Begin: 1976/3rd End: 1978/2nd Qtr
21 EVALUATE THERMAL $200,000 REGENERATION OF NOVEL ION EXCHANGE MATERIAL Contract Begin: 1976/3rd End: 1978/2nd Qtr
C.a. Decision Sufficiency Of Data
24 PREPARE MANUAL OF $100,000 PRACTICE FOR ION EXCHANGE Contract Begin: 1978/3rd End: 1979/2nd Qtr
D. DISTILLATION
02, 08 LAB & PILOT STUDIES OF DISTILLATION
25 DEMONSTRATE $450,00 DISTILLATION OF WASTEWATER Demonstration Grant Begin: 1976 3rd End: 1980/2nd Qtr
D.a. Decision Sufficiency Of Data
28 PREPARE MANUAL OF $100,000 PRACTICE FOR DISTILLATION Contract Begin: 1980/3rd End: 1981/2nd Qtr
PRE-FY 74
FY 74 75
FY 76
FY 77 78
FY 79
POST-FY 79

CHART VI (continued)

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TREATMENT AND DISPOSAL OF SLUDGE FROM MUNICIPAL WASTEWATER PLANTS IN THE UNITED STATES

DR. JOSEPH B. FARRELL, CHIEF
ULTIMATE DISPOSAL SECTION
TREATMENT PROCESS DEVELOPMENT BRANCH
ADVANCED WASTE TREATMENT RESEARCH LABORATORY
NATIONAL ENVIRONMENTAL RESEARCH CENTER
ENVIRONMENTAL PROTECTION AGENCY
CINCINNATI, OHIO

PRESENTED AT

THIRD U.S./JAPAN CONFERENCE ON SEWAGE TREATMENT TECHNOLOGY
TOKYO, JAPAN

FEBRUARY 1974

TREATMENT AND DISPOSAL OF SLUDGE FROM MUNICIPAL WASTEWATER PLANTS IN THE UNITED STATES *

Most communities in the United States have primary treatment and secondary treatment is a national goal. The quantities of sludge which must be handled are large (Table 1). We expect a large increase in the amount of secondary and chemical sludges. As Table 1 shows, most of our sludge is disposed to the land. A substantial proportion is incinerated and the ash is disposed to land. Ocean disposal of sludge is expected to be greatly reduced in the future.

Composition of Sludge

Municipal sludge is chiefly composed of paper fiber, human wastes, food wastes, a proportion of industrial wastes, and, when stormwater is included in the sewers, soil and dirt from roads. Raw sludges may contain 70 to 85 percent volatile solids (30 to 15% ash) and digested sludge 50 to 65 percent volatile solids (50 to 35% ash). Sludge from a community contains trace amounts of hazardous materials. Typical values are presented in Tables 2 and 3. The range of values can be extremely high. It is important that communities periodically analyze their wastewater sludge to determine whether hazardous levels of certain contaminants will place restraints on their method of disposal.

Ocean Disposal**

The United States Council for Environmental Quality has spoken out for an eventual cessation of the disposal of sewage sludge by barging to disposal sites in the ocean. Most barging is conducted on our East Coast where the sludge is dumped in the relatively shallow waters of the Continental Shelf. Regulations have been published in the Federal Register(1) and the Code of Federal Regulations(2) listing maximum concentrations of certain metals and compounds which should not be exceeded in material being dumped. For example, it is stated that mercury in the solid phase should not exceed 0.75 mg/kg. Virtually any municipal sludge contains at least several times this concentration of mercury. There is pressure to raise these levels but it is unlikely that changes will be made. The United States Environmental Protection Agency is issuing permits for dumping of sewage sludges which are described as interim permits. Municipalities are required to prepare a plan for meeting the desired concentrations and reapply annually for a permit. It is anticipated that municipalities will find other disposal methods less costly than meeting the ocean disposal regulations. The EPA has prepared a report describing the progress of the ocean dumping program(3).

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* Presented at Third U.S./Japan Conference on Sewage Treatment Technology, Tokyo, Japan, February 13, 1974.

** The author of this paper has not dealt intimately with ocean disposal and is not an authority on EPA's Ocean Dumping Policy. The discussion is presented for general guidance only.

For years, cities in California have discharged digested sludges by pipeline to outfalls into the ocean. There is essentially no continental shelf on the Western Coast of the U. S. Nevertheless, California has a policy which will probably prohibit ocean disposal of sludges in the foreseeable future.

Stabilization

Anaerobic Digestion -- There has been a renewed interest in anaerobic digestion because of the energy shortage. Most of the interest is related to agricultural and animal wastes and municipal solid wastes. The quantity of recoverable methane from sewage sludge digestion is too low to have an impact on the national needs. However, it can help the economics of individual plants. Most wastewater treatment plants which utilize anaerobic digestion waste the surplus gas over their needs for sludge heating. Los Angeles is a notable exception. Both the city and the county sell their excess gas production to nearby utilities. There will probably be more efforts made to utilize excess gas for in-plant power generation and fuel needs. Its impact will be small.

Aerobic Digestion -- Aerobic digestion is practiced at many small plants where it is often followed by land disposal of liquid sludge. We have supported research work aimed at improving plant scale operation and establishing design information(4). We have also supported work on thermophilic aerobic digestion (about 60° C) utilizing oxygen. This work, although in its early stages, has been encouraging. It has the potential of faster digestion (hence smaller equipment) and will produce a sludge containing no pathogens.

Chlorine Stabilization -- Stabilization of sludge by treating it with high concentrations of chlorine (about 2000 mg/l) produces a stable and sterile sludge. This process is being used at numerous small plants. We are concerned that the drainings from this sludge may contain toxic chlorinated compounds. We recommend that all drainings be recirculated to the incoming sewage.

Lime Stabilization -- Our studies(5) show that addition of lime to liquid sludge will stabilize the sludge for a sufficiently long time to permit nuisance-free disposal. We now recommend adding lime to liquid sludge to raise pH to 12. This will require less than 0.15 kg $Ca(OH)_2$ per kg of sludge (dry solids basis). The sludge and lime are preferably mixed by air, which removes the odor of ammonia. Pathogenic bacteria are eliminated. The sludge can be dried on sand beds, disposed to a landfill, or spread in liquid form on farm land. It should not be discharged to a lagoon or left in deep piles on the surface, because pH will eventually drop and putrefaction can occur.

Microbiological Destruction

The disposal of digested sludge to landfills should represent no serious microbiological problem. Solid waste is substantially higher in pathogenic activity than digested sludges.

When there is any large-scale disposal of digested sludge to agricultural land, the question of microbial contamination is generally brought forth. If proper precautions are taken, there is no evidence of an undue hazard. In some countries, however, pasteurization is required during the summer. Recently, a plant utilizing nuclear isotopes to irradiate sludge went on stream in Geiselbullach, Germany. It is believed that their concern is chiefly worm eggs and cysts. It is possible that many local jurisdictions will lean in the direction of extreme caution and will require the equivalent of pasteurization for large scale agricultural utilization of sludge.

Sludge Conditioning

Chemicals -- The use of inorganic chemicals such as ferric chloride and lime continues to lose ground. Anionic, cationic, and nonionic polymers are used in increasing amounts to condition sludges for the elutriation process, for gravity and air flotation thickening, and for dewatering. The EPA has not found it necessary to stimulate development in this field by grants and contracts. Several very competent chemical companies are engaged in research and development of polymer conditioning agents and are heavily committed to developing this market.

The use of polymers has made it possible to continue the use of elutriation as a preliminary step for dewatering digested sludge. When primary and activated sludge are digested together, elutriation often washes out a large proportion of fines. These fines are returned to process where they cause deterioration in effluent quality. EPA considered making elutriation ineligible for construction grant funds. However, the use of polymers has improved elutriation performance sufficiently that it is still possible to get approval for construction grants for elutriation facilities.

Thermal Conditioning -- The use of heat to condition sludge for dewatering has seen rapid growth in recent years. For example, from 1970 through about the first six months of 1973, Zimpro Division of Sterling Drug installed or has under construction its low pressure sludge oxidation system for a population equivalent of about 10 million people. Other companies such as Envirotech (Porteous Heat Treatment Process) have had a similar experience.

Experience of communities with sludge conditioning equipment is mixed. Some communities are very pleased with the process whereas others have had many difficulties. The usual complaints are failure of equipment (often related to stones and metal in the sludge), odor, excessive cost, and high BOD load and color in the supernatant. Limited bench-scale studies at the Taft Center showed no substantial change in performance of the activated sludge process when heat treatment supernatant was included in the incoming primary effluent (COD load was held constant) except for a persistent yellow color in the final effluent. Zimpro and others(6) have found that activated carbon removes this color effectively.

The heat treatment of sludge has many vigorous advocates and equally vigorous opponents. It is this writer's opinion that, barring the unlikely discovery that heat treatment causes formation of materials extremely hazardous to health, heat conditioning of sludge is a viable wastewater treatment process. When its use is considered, its impact on the cost of the entire processing sequence should be considered. The bad experiences at some plants indicate that manufacturer's recommendations of operating and maintenance costs and off-stream time for repairs should be adjusted upwards when cost estimates are made.

Dewatering

High Solids -- A new factor--the solids content of the dewatered sludge--is becoming important in dewatering technology in the United States. In the higher rainfall areas of the United States, sludge which is disposed to a landfill should carry with it a minimum amount of water. If the dewatered sludge is relatively dry, it will be easier to handle and will contribute less to leachate than a wetter sludge. Several states have put restrictions on solids content that range from general statements to the effect that the sludge shall be easily handled by conventional landfill equipment to a specific statement by one state that the sludge shall contain at least 50 percent dry solids.

A second consideration which makes solid content important is the energy shortage. If sludge is being incinerated, supplementary fuel will be needed during burning unless the solids content is above about 30 percent. With the cost of desirable fuels such as gas and fuel oil expected to double in price and often be unavailable, there will be heavy pressure to produce a high solids dewatered sludge.

Vacuum filters and centrifuges rarely produce cake solids greater than about 22 percent solids when dewatering a mixture of primary and activated sludges. Pressure filters can produce sludge cake in the desired moisture range. Up until now, pressure filtration has been very slow in penetrating the United States market, chiefly because the capital cost is so high.

We are clearly in need of innovative means to produce dewatered sludge with solids content in excess of 30 percent solids. The development of equipment which can remove additional water from sludge cake after it leaves a centrifuge or vacuum filter is especially attractive.

Belt Filters -- Two primary types of belt filters have become available in the United States: the Carter Belt-Filter press and the Westinghouse Capillary Suction Dewatering System.

The Carter Belt-Filter press is a German development. It comprises three dewatering zones: initial draining, pressing, and shearing. Sludge is contained between two woven belts (0.2 to 1.5 mm openings). Rollers are carefully positioned to apply pressing and shearing action to the sludge. There were approximately 1,000 of these units installed in Germany by 1971.

The Westinghouse (Infilco Division) Capillary Suction Dewatering System has been developed with partial assistance from EPA. It utilizes a porous belt which removes water from sludge by "capillary suction." Near the end of the belt travel, a smooth roll, driven at the same linear speed as the belt presses against it. The pressure forces additional water from the sludge into the capillary belt. The sludge cake transfers to the smooth roll where it is scraped off and conveyed away. This device works well on sludges which are difficult to dewater. Polymer demands are low and production rate is high. A demonstration unit is being evaluated at St. Charles, Illinois.

Both of these devices offer possibilities of slightly higher sludge solids than solid-bowl centrifuges or vacuum filters. They are simple to operate and should have low maintenance costs.

Top-Feed Filter -- Difficulty is often encountered in conventional vacuum filtration with pickup from the sludge pan, and with cake release. Experiments have been conducted at Milwaukee with the aid of an EPA grant with a vacuum filter in which the sludge pan was moved from the usual 4:00 o'clock to 8:00 o'clock position on the drum up to the 8:00 o'clock to 11:00 o'clock position. Difficulty encountered with sealing the pan was overcome. Sludge pickup was excellent. Cake release also was excellent. The cake release point was at the 8:00 o'clock position. The force of gravity was an important aid in removing cake from the filter cloth. Cake solids and filter yield were increased. A grant to demonstrate the top-feed filter on a large scale has been made to Milwaukee.

Centrifugation -- Los Angeles County has demonstrated the utility of a centrifuge of the vertical basket-type (solid wall basket) for removing fine solids from sludge with a high recovery. The minimum cost method for Los Angeles to improve the performance of their solid-bowl centrifuges and produce a high solids cake was to process the centrate through the basket-type centrifuge. The basket-type centrifuge can give high recovery of solids at low conditioning chemical cost. However, cake solids is lower than obtained with the solid-bowl centrifuge.

Disposal

Incineration -- An EPA Task Force conducted a series of tests on sewage sludge incinerators and reported its results on air pollution and metals concentration in particulates in a Task Force Report(7). The recommendations of the Task Force are found in another publication(8). On the basis of the above-mentioned tests, proposed standards of performance for new stationary sources have been published(9). The standards specify only that the exhaust gas contain less than 70 mg/Nm^3 (0.031 grain/dry scf) of particulate matter. This writer believes that in applying this standard, wet scrubbing will be required, probably with a venturi-type scrubber or similar (ca. 100 cm water pressure drop). Some manufacturers maintain that a low pressure drop scrubber (ca. 15 cm water) will clean the gases adequately and are attempting to change this requirement.

The Task Force on incineration recommended that afterburners be used to insure that polychlorinated biphenyls and other organic materials are destroyed. The suggested conditions were 1600° F (870° C) for 2.0 seconds or a combustion condition that accomplishes the desired destruction. Tests by manufacturers indicate that an afterburner temperature of 1100° F (590° C) and about 0.5 second accomplishes virtually complete destruction of these materials when they are incinerated with sewage sludge in a multiple hearth incinerator.

Incinerators can be made to meet virtually [illegible]e air pollution requirement without becoming extremely costly. [illegible]s a report of an incinerator being permitted in the San Francisco Air Pollution Control District because it would produce less pollution than would the trucks needed to transport the sludge cake to a suitable sanitary landfill.

Landfill -- Proposed guidelines have been published for land disposal of solid wastes(10). These guidelines are not a binding obligation. The disposal of digested sewage sludges is permitted in sanitary landfills where there is provision for handling these sludges. The sludges must be digested and must contain no "free" water ("free" water is not defined).

Much sewage sludge is disposed by municipalities in what might be called "private" landfills, but which are often nothing more than dumping grounds. It is our intent to promptly commence the development of an environmentally acceptable procedure for disposal of sludge from small wastewater plants to private landfills.

Land Spreading -- A critical review of land spreading practice in the United States has been made(11) and will be published soon. No detailed discussion will be presented here. Up until recently, most of our attention has been given to nitrogen as the factor that limits the rate of application of sludges to the land. For communities with a high proportion of industrial wastes, the presence of certain metals which can be taken up by growing crops will prevent the use of land spreading or limit loading rates to uneconomical levels. The loading levels of metals which are eventually permitted in land spreading applications will decide whether this conserving use of sludge will be an economically viable alternative to other disposal methods.

J. B. Farrell
January 31, 1974

LITERATURE CITED

1. Federal Register, 38, No. 198, Part II, "Ocean Dumping, Final Regulations and Criteria," Oct. 15, 1973.

2. Code of Federal Regulations, 40 CFR, 220-227.

3. "Annual Report on the Administration of the Ocean Dumping Program, Fiscal Year 1973," pub. U. S. EPA.

4. Cohen, D. B., and J. L. Puntenny, "Metro Denver's Experiences with Large Scale Aerobic Digestion of Waste Activated Sludge," presented at 47th Annual Conf. WPCF, Cleveland, Ohio, Oct. 4, 1973.

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TABLE 1: MUNICIPAL SLUDGES FOR DISPOSAL

QUANTITIES

SLUDGE TYPE	1972		1985	
	POPULATION (MILL.)	TON/YR.	POPULATION (MILL.)	TON/YR.
Primary (0.12 lb/cap-da)	145	3,170,000	170	3,720,000
Secondary (0.08 lb/cap-da)	101	1,480,000	170	2,480,000
Chemical (0.05 lb/cap-da)	10	91,000	50	455,000

DISPOSAL METHODS	1972	1985
	(% of Population)	
Landfill	40	40
Utilized on Land	20	30
Incineration	25	30
Ocean (Dumping and Outfalls)	15	0

TABLE 2

CONCENTRATION OF METALS IN SLUDGE -- SEVEN U. S. LOCATIONS

Metal	Median (mg/kg)	Range (mg/kg)
Beryllium	0	N.D. *
Cadmium	200	N.D. to 800
Chromium	1,800	400 - 5,900
Copper	1,700	900 to 6,000
Mercury **	4.5	3.0 to 5.5
Lead	2,800	800 to 6,900
Zinc	1,600	400 to 8,400

* N.D. = not detected

** Only 3 sludges analyzed

TABLE 3

PESTICIDE AND PCB LEVELS IN SLUDGES

Compound	Number of Sites Sampled	Median (mg/kg)	Range (mg/kg)
Aldrin	3	N.D. *	16 (in one sample only)
Dieldrin	5	0.3	0.08 to 2.0
Chlordane	5	18.4	3.0 to 32
DDD	5	0.2	N.D. to 0.5
DDT	5	0.2	N.D. to 1.1
PCB	10	2.8	N.D. to 105

* N.D. = not detected

EXPERIENCES WITH SLUDGE HANDLING IN TEXAS

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TOKYO, JAPAN

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EXPERIENCES WITH SLUDGE HANDLING IN TEXAS

Texas, located in the southwestern part of the United States, is a large state covering 267,339 square miles (692,408 square kilometers.)(1) Some areas are highly urbanized, others are barely inhabited. For example, Dallas County has a population density of about 1500 persons per square mile (580 persons/square kilometer); while Kenedy County has a population density of about 0.5 person per square mile (0.2 person/square kilometer.)(1) As a consequence of the varying degrees of urbanization, Texas has both large sewage treatment plants and very small sewage treatment plants--ranging in treatment capacities from roughly 100 MGD (380,000 m^3/day) to 2000 gallons per day (7.6 m^3/day.) This same factor dictates that plants are located in both highly populated areas and very remote rural areas. The climate in Texas varies from humid in the East to arid in the West--rainfall variations from roughly 50 inches per year (127 cm/year) to less than 8 inches per year (20 cm/year.) These diversities lead to a host of sludge handling and disposal techniques.

The predominate sludge disposal technique utilized by small plants in rural areas in Texas is anaerobic digestion with dewatering on open beds and subsequent land application. This technique is satisfactory and will continue in popularity.

At small plants in urban areas, the trend is away from anaerobic digestion and toward aerobic digestion. This trend is created by two considerations: (1) the extensive use of contact stabilization process treatment plants and the amenability of aerobic digestion with this process, and (2) nuisance problems associated with handling anaerobic sludges.

Where small plants utilizing aerobic digestion are constructed within economical hauling range of a large plant with sludge handling capability, drying beds are not constructed--the excess sludge produced being transported to a sludge plant for dewatering and disposal. Since many of these plants depend upon contract hauling of the sludge with the attendant unreliability of transportation, difficulties have been experienced with sludge buildup within the treatment process and subsequent deterioration of effluent quality. This is, of course, not inherent in the scheme--rather, a defect in the management and implementation.

The City of Houston employs a novel technique for transporting excess activated sludge from plants with no dewatering facilities to a sludge processing plant. The City is located on the coastal plain which has a flat topography. Because of topography and other factors, the City has thirteen permanent sewage treatment plants and other temporary plants. Only two plants, the two largest, are equipped with sludge dewatering facilities. The excess activated sludge from some of the plants is pumped via pressure conduit to the nearest sanitary sewer which flows to the major plants equipped with sludge dewatering facilities. No problems have arisen as a result of this transportation technique.

Where sludge facilities are not within economical or practical hauling distance, small treatment plants utilizing aerobic digestion are equipped with sludge drying beds. Difficulties have been encountered in dewatering such sludges on drying beds--primarily the blinding of the bed. In coping with this problem, operators have learned to draw sludge onto the beds very slowly. This permits the supernatant to flow over the settled and settling sludge and drain through the sand bed ahead of the encroaching sludge. At some plants, arrangements have been made to fill the bed, allow the sludge to settle, and decant the supernatant. The addition of polymers to the sludge as it is being drawn is reported to help overcome the blinding of the bed. The polymer is added to the aerobic digester about 12 hours before drawing.(4) The addition of polymers to the sludge as it is being drawn is reported to help overcome this problem.

At large plants in Texas, various sludge disposal techniques are employed.

The City of Houston, at the two major facilities previously mentioned, utilizes the activated sludge process without primary sedimentation. The excess sludge is chemically conditioned with ferric chloride (75 lbs. per ton of dry solids - 3.75%), dewatered on rotary vacuum filters, flash dried and, subsequently, sold as a fertilizer under the name of "Hou-Actinite." The sludge to the filters typically average around 4% solids, the filter cake about 15%, and the completely dried processed sludge about 95.7%.(2) The fertilizer typically has a moisture content of around 5%; ash content of 35%; nitrogen content of 5%; and available phosphoric acid, 4%.(2) The revenue to the City in the first six months of 1972 from fertilizer sales was reported to be $21 per ton (907 kg.)(2)

The City of Houston is presently building an additional 30-ton-per-day (27,000 kg/day) sludge processing plant at its Alameda Plaza sewage treatment plant. This plant is to employ pressure filtration in lieu of vacuum filtration ahead of flash drying. In order to produce a filterable sludge, the sludge must be conditioned with both ferric chloride and lime.(3)

It is estimated that the pressure filter cake will have a solids content of about 35% as contrasted to vacuum filter cake with a solids content of 15%.(3) The increased solids content permits the drying step to be accomplished with smaller flash dryers and a decreased fuel consumption. The savings in drying cost by using the pressure filters dictated the design change from vacuum filters. One disadvantage of the design change has been to lower the value of the fertilizer produced. This is brought about by the dilution of the nutrient value of the fertilizer by lime used in sludge conditioning and diatomaceous earth used to precoat the filters. It is estimated that the ash content will be increased to 49% and the nitrogen content lowered to about 4%.(3) The pressure filter-flash dryer system is estimated to save the City about 2.5 million dollars over the next nine years.(3)

The Gulf Coast Waste Disposal Authority operates a large industrial waste treatment plant in the Houston area known as the Champion plant. The plant treats paper mill waste with a small amount of other waste, primarily petroleum refining waste. The plant employs the activated sludge process and has a treatment capacity of 44 mgd (166,500 m^3/day).

Centrifuges are used to dewater excess activated sludge. The dewatered sludge is conveyed to barges moored in the ship channel for transportation to ultimate disposal in lagoons constructed in the coastal wetlands. The sludge is conditioned with polymer prior to centrifugation. A 20% solids cake is generally produced.(4) The centrifuges in this operation experience unusually high levels of abrasion wear. This wear is attributed to the lime and titanium content of the sludge derived from the paper-making process.(4) Because of the centrifuge problem and the likelihood that the lagooning operation in the coastal wetlands will be declared unacceptable by the regulatory agencies, the Authority is looking toward other sludge disposal techniques. They have tentatively decided upon pressure filtration and land filling. The primary factors influencing the decision to go to pressure filters over vacuum filters, (2) chemical conditioning cost associated with vacuum filters, and (3) the dryer cake produced by the pressure filters.(4) The dryer cake facilitates subsequent drying and/or incineration should these steps be ultimately required, and also permits transportation of the cake in open dump trucks without dripping liquids on the public roadways should an inland landfill site be chosen.

The City of Austin Govalle sewage treatment plant utilizes lagooning to dispose of its excess activated sludge. The plant is a 40 mgd (151,400 m^3/day) contact stabilization plant. The lagoons are constructed above grade and the only liquids they receive are excess activated sludge, the rain that falls on their surface, and river water pumped into them to maintain a constant water level. River water pumpage is necessary to keep a constant water level since the excess activated sludge is not sufficient to make up for evaporation losses. The ponds cover some 191 acres (77.3 hectares) and are so constructed that they have a uniform depth of 8 feet *2.43 m) and smooth sides with a 1/4 slope. A paddle boat is provided to break up scum which sometimes forms. This system has worked extremely well and no problems have been encountered with nuisance conditions. In fact, the ponds are one of the favorite areas of bird watchers in the winter when the ponds are visited by great numbers of water fowl and shore birds.

The City of San Antonio also employs sludge lagooning at its Rilling Road plant. This plant is a 94 mgd (355,790 m^3/day) conventional activated sludge plant. Only the excess activated sludge is lagooned--the primary sludge is subjected to anaerobic digestion and drying on open beds. The sludge lagoon is known as Mitchell Lake. It is actually an articicial impoundment covering some 850 acres (344 hectares) with a watershed area of 5000 acres (2023 hectares). The lake has been in continuous service as a sludge lagoon for 70 years, and it is interesting to note that the sludge accumulation over the lake bottom varies from about 3 feet (.91 m) in the upper and to about 1 foot (.305 m) in the lower end. Nuisance conditions frequently exist at this lagoon. These conditions are created by extensive shallows which are frequently exposed to the atmosphere by the fluctuating water level in the lake. The water level in the lake is allowed to fluctuate, getting low in dry weather and overflowing in periods of adequate rainfall. Due to the public dissatisfaction with Mitchell Lake, the City is presently considering alternate sludge disposal techniques.

It is our expectation that large plants in Texas in the more arid portions of the State will continue to utilize sludge lagooning where feasible. In heavily populated areas where ultimate disposal is fertilizer production or long distance landfill, particularly where the landfill must be reached via public roads, we expect a trend away from vacuum filters to pressure filters. Where ultimate disposal is nearby landfill, vacuum filters will most likely continue to be the most viable method of dewatering.

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PHYSICAL-CHEMICAL NITROGEN REMOVAL WASTEWATER TREATMENT

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INTRODUCTION

There are three basic physical-chemical nitrogen-removal techniques available for application today. These three processes are

- Ammonia stripping (ch. I)
- Selective ion exchange (ch. II)
- Breakpoint chlorination (ch. III)

All of these approaches have the advantage that they are based on the removal of nitrogen in ammonia form, which eliminates the costs of converting the ammonia to nitrate in the biologic-treatment step. They also have the advantages that they are unaffected by toxic compounds that can disrupt the performance of a biologic nitrogen-removal system, they are predictable in performance, and the space requirements for the treatment units are less than for biologic-treatment units.

The advantages and disadvantages of each of these physical-chemical processes are discussed in detail and the processes are compared in the chapters that follow. Discussion of these processes includes application at the following facilities, either in existence or under design:

- South Lake Tahoe, Calif.
- Orange County, Calif.
- Windhoek, South Africa
- Blue Plains, D.C.
- Upper Occoquan Sewage Authority, Va.
- Rosemount, Minn.
- North Lake Tahoe, Calif.
- Montgomery County, Md.
- Cortland, N.Y.

Chapter I

AMMONIA STRIPPING

The only nitrogen-removal process that actually has been used on a plant scale in wastewater treatment is ammonia stripping. This process has been in use for ammonia nitrogen at the South Lake Tahoe plant for about 4 years. Both the advantages and limitations of this process have been clearly demonstrated.

The ammonia-stripping process itself consists of

- Raising the pH of the water to values in the range of 10.8 to 11.5, generally with the lime used for phosphorus removal
- Formation and re-formation of water droplets in a stripping tower
- Providing air-water contact and droplet agitation by circulation of large quantities of air through the tower

The towers used for ammonia stripping closely resemble conventional cooling towers.

Questions are sometimes raised concerning the fate of ammonia discharged to the atmosphere. Are we merely converting a water-pollution problem to an air-pollution problem? Does the ammonia stripped from the wastewater cause an air-pollution problem or find its way back to the receiving stream owing to scavenging by precipitation?

The concentration of ammonia in the stripping-tower discharge is only about 6 mg/m^3 for domestic wastewaters (at an air flow of 500 ft^3/gal and at an ammonia concentration of 23 mg/l in the tower influent). As the odor threshold of ammonia is 35 mg/m^3, the process does not present a pollution problem in this respect. The ammonia discharged to the atmosphere is a stable material that is not oxidized to nitrogen oxides in the atmosphere. The natural production and release of ammonia as part of the natural nitrogen cycle is about 50 billion tons per year. Roughly 99.9 percent of the atmosphere's ammonia concentration is produced by natural biological processes.[1] There is a large turnover of ammonia in the atmosphere, with the total ammonia content being displaced once a week on the average. Ammonia is returned to the earth through gaseous deposition (60 percent), aerosol deposition (22 percent), and precipitation (18 percent). Ammonia is not considered an air pollutant because there are no known public health implications, and because it is a natural constituent of the atmosphere derived almost entirely from natural sources. For example, a single cow releases as much nitrogen to the atmosphere in feces and urine as 12 people would contribute if all of their ammonia production were stripped to the atmosphere.

There are no standards in the United States for ammonia concentrations in the atmosphere. Some foreign standards[1] have been established.

- Czechoslovakia, 100 mg/m^3 (24 hours)
- U.S.S.R., 200 mg/m^3 (24 hours)
- Ontario, Canada, 3,500 mg/m^3 (30 minutes)

All of these standards are far above the 6 mg/m^3 that will occur right at the tower discharge. The process cannot be dismissed from consideration because of air pollution.

A remaining question is the fate of the ammonia discharge to the air. Is it likely to find its way into the receiving stream by being scavenged from the atmosphere by precipitation?

Ammonia may be washed from air by rainfall, but not by snowfall. The natural background concentration of ammonia in the atmosphere is 5-7 ppb. In rainfall the natural background ranges from 0.01 to 1 mg/l, with the most frequently reported values of 0.1 to 0.2 mg/l. The amount of ammonia in rainfall is related directly to the concentration of ammonia in the atmosphere. Thus, an increase in the ammonia in rainfall wuuld occur only in that area where the stripping-tower discharge increases the natural background ammonia concentration in the atmosphere.

Calculations for the ammonia washout in a rainfall rate of 3 mm/h (0.12 in./h) have been made for the Orange County, Calif., project. The ammonia concentrations of ammonia in the rainfall would approach natural background levels within 16,000 feet of the tower. Of course, the ammonia discharge during dry periods diffuses into the atmosphere quickly so that the background concentration and resulting washout rate of ammonia at greater distances from the tower are not affected during a subsequent storm. The ultimate fate of the ammonia that is washed out by rainfall within the 16,000-foot downwind distance depends on the nature of the surface upon which it falls. Most soils will retain the ammonia. That portion which lands on paved areas or directly on a stream surface will appear in the runoff from that area. Even though a portion of the ammonia washed out by precipitation will find its way into surface runoff, the net discharge of ammonia to the aquatic environment in the vicinity of the plant would be very substantially reduced.

One of the great advantages of this method of nitrogen removal is its extreme simplicity. Water is merely pumped to the top of the tower at a high pH, air is drawn through the fill, and the ammonia is stripped from the water droplets. The only control required is the proper pH in the influent water. This simplicity of operation also enhances the reliability of the process.

Several factors affect the efficiency of the ammonia-stripping process.

- Type of stripping unit
- pH
- Temperature
- Loading rate
- Scale of deposition

There are three basic types of stripping units now being used in full-scale applications.

- Countercurrent towers
- Crossflow towers
- Stripping ponds

Countercurrent towers (the entire airflow enters at the bottom of the tower while the water enters the top of the tower and falls to the bottom) have been found to be the most efficient. In the crossflow towers, the air is pulled into the tower through its sides throughout the height of the packing. This type of tower has been found to be more prone to scaling problems. The stripping-pond approach will be discussed in more detail later.

The pH of the water has a major effect on the efficiency of the process. The pH must be raised to the point that all of the ammonium ion is converted to ammonia gas. The pH required varies somewhat with temperature,[2] but is generally about 11.0.

Another critical factor is the air temperature. The water temperature has less effect on performance because the water temperature reaches equilibrium with the air temperature in the top few inches of the stripping tower. The efficiency of the process decreases as the temperature decreases. For example, at 20° C 90 percent removal of ammonia is typically achieved. At 10° C, the maximum removal efficiency drops to about 75 percent. When air temperatures reach freezing, the tower operation must generally be shut down owing to icing problems.

The hydraulic loading rate of the tower is also an important factor. This rate typically is expressed in terms of gallons per minute applied to each square foot of the plan area of the tower packing. When the hydraulic loading rates become too high, good droplet formation is disrupted and the water begins to flow in sheets. Tower loading rates of 2 gal/min/ft^2 have been shown to be compatible with optimum tower performance.[2] It is critical that the water and air be uniformly distributed over the tower area.

Another factor that may have an adverse effect on tower efficiency is scaling of the tower packing resulting from deposition of calcium carbonate from the unstable, high-pH water flowing through the tower. The original crossflow tower at the South Lake Tahoe plant has suffered a severe scaling problem. The severity of the scaling problem was not anticipated from the pilot studies in which a countercurrent tower was used. As a result, the full-scale crossflow-tower packing was not designed with access for scale removal in mind. Thus, portions of the tower packing are inaccessible for cleaning. Those portions that were accessible were readily cleaned by high-pressure hosing. The potential scaling problem must be recognized in design. The use of countercurrent towers and design of the packing with access for cleaning can adequately combat this problem.

An example of design for scale control is the 15-mgd tower now under construction at the Orange County, Calif., Water District plant (fig. I-1). There the tower packing has been designed to be readily removable for cleaning as a precaution against scaling problems, although no significant scaling problem has been observed in several months of pilot tests at Orange County.[3] Scaling has also been reported not to be a significant problem at the Windhoek, South Africa, plant where only a soft, easily removed scale was encountered.[4] On the other hand, tests at the Blue Plains pilot plant encountered a hard scale that was extremely difficult to remove.[5] The hardness of the scale at Blue Plains was affected by operating pH, with a harder scale forming at pH 11.5 than at pH 10.8.

Typical design criteria are

- Hydraulic loading, 1 to 3 gal/min/ft^2
- Air-to-water ratio, 300 to 500 ft^3/min per gal/min
- Air-pressure drop, 0.5 to 1.25 inches water
- Fan-tip speed, 9,000 to 12,000 ft/min
- Fan-motor speed, 1 or 2 speed
- Packing depth, 20 to 25 feet

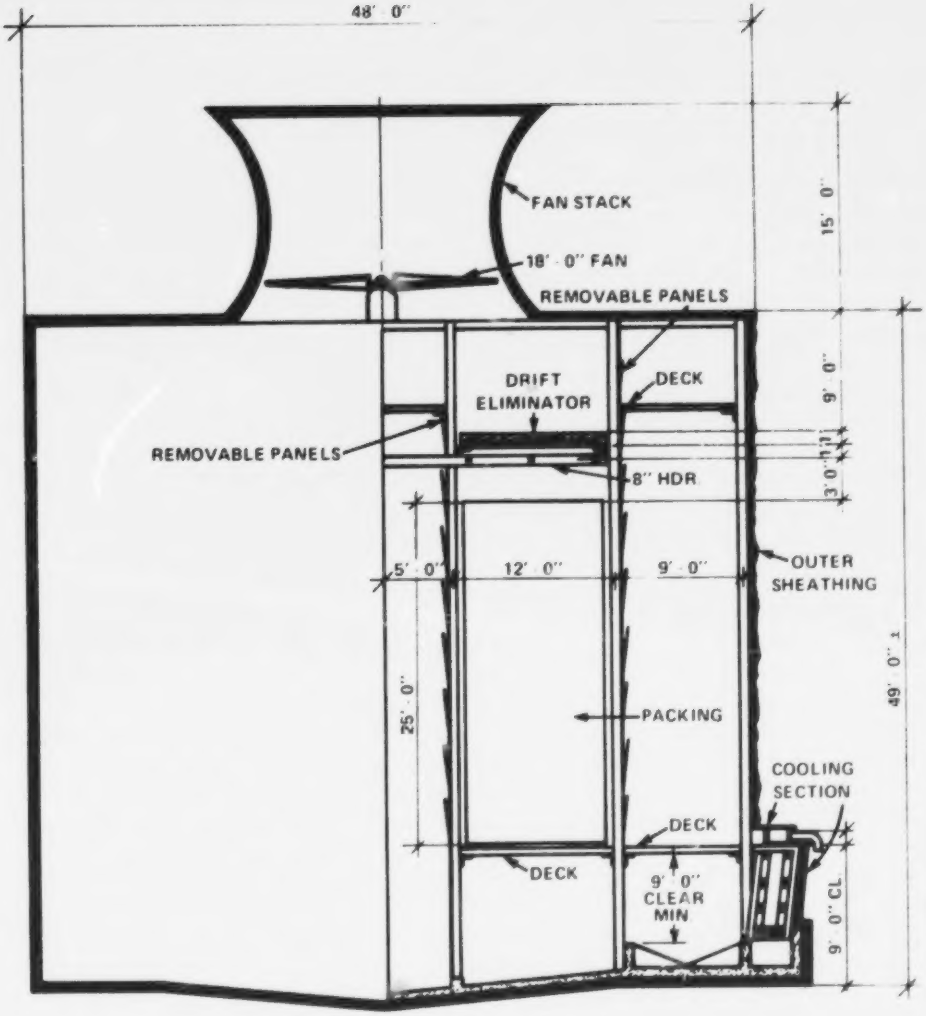

Figure I-1. Ammonia-stripping tower design, Orange County, Calif.

- Packing spacing, 2 to 4 inches horizontal and vertical
- Packing material, wood, plastic (½-in. PVC pipe being used at Orange County)

A curve for estimating the costs of the ammonia-stripping process for various-size plants is presented in figure I-2. This curve is based on a loading rate of 2 gal/min/ft^2. Because some applications may require ammonia removal only during warm weather months, operating costs are shown for both 6-month and 12-month operation.

The South Tahoe system is being modified to reduce the impact of temperature and scaling limitations encountered at this plant.[6] Basically, the modified process will consist of three steps (see figs. I-3, I-4, and I-5).

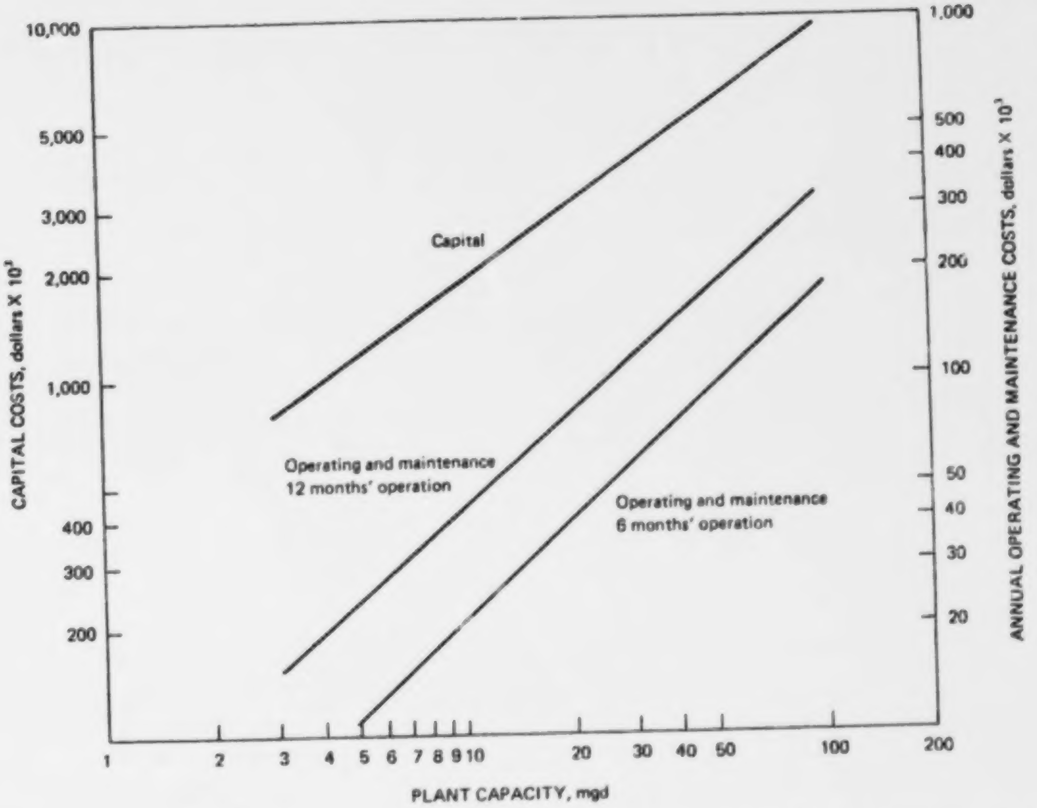

Figure I-2. Ammonia-stripping costs. (EPA STP Index = 200; includes engineering, legal, administrative, construction financing, and contingencies.)

- Holding in high-pH, surface-agitated ponds
- Stripping in a modified, crossflow forced-draft tower through air sprays installed in the tower
- Breakpoint chlorination

This system was inspired by observations in Israel of ammonia nitrogen losses from high-pH holding ponds.[7]

Pilot tests at South Tahoe indicated that the release of ammonia from high-pH ponds could be accelerated by agitation of the pond surface. In the modified Tahoe system, the high-pH effluent from the lime clarification process will flow to holding ponds. Holding pond detention times of 7-18 hours will be used in the modified South Tahoe plant. The pond contents will be agitated and recycled 4-13 times by pumping the pond contents through vertical spray nozzles into the air above the ponds. At least 37 percent ammonia removal is aniticipated, even in cold weather conditions, in the ponds. The pond contents then will be sprayed into the forced-draft tower. The packing will be removed from the tower and the entire area of the tower will be equipped with water sprays. At least 42 percent removal of the ammonia in the pond effluent is anticipated, based on pilot tests, from this added spraying in cold weather, which will include recycling of the pond effluent through the tower to achieve 2-5 spraying cycles. The ammonia escaping this process then

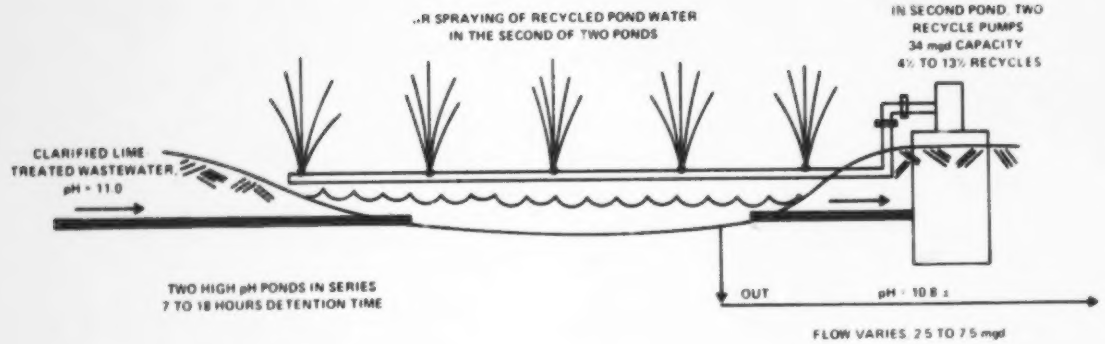
..R SPRAYING OF RECYCLED POND WATER
IN THE SECOND OF TWO PONDS
IN SECOND POND, TWO
RECYCLE PUMPS
34 mgd CAPACITY
4% TO 13% RECYCLES
CLARIFIED LIME
TREATED WASTEWATER,
pH = 11.0
TWO HIGH pH PONDS IN SERIES
7 TO 18 HOURS DETENTION TIME
OUT
pH · 10.8 ±
FLOW VARIES 2.5 TO 7.5 mgd

Figure I-3. Proposed new and modified ammonia nitrogen removal processes, South Lake Tahoe: New high-pH flow-equalization ponds.

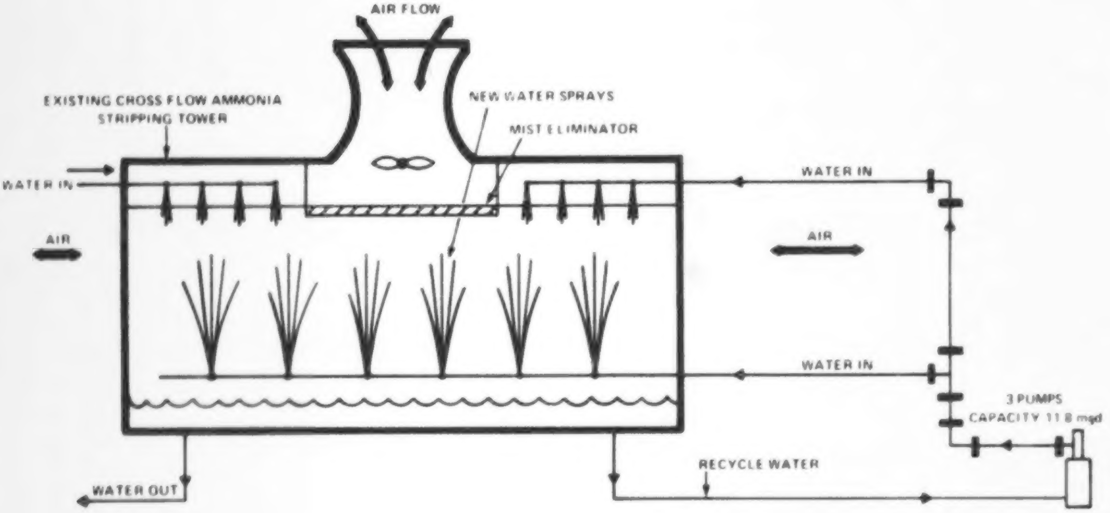

Figure I-4. Proposed new and modified ammonia nitrogen removal processes, South Lake Tahoe: Existing stripping tower modified with new sprays.

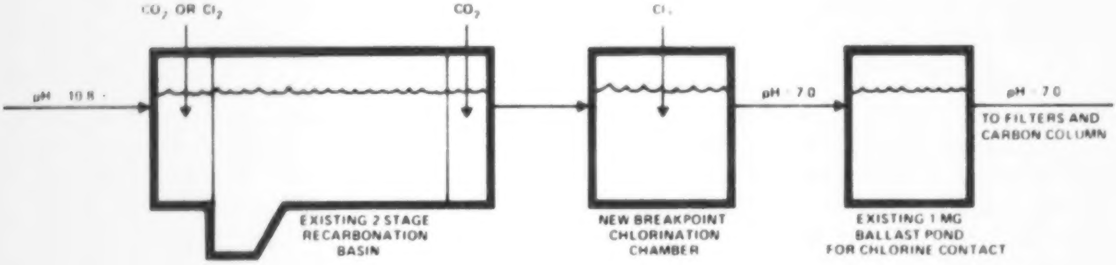

Figure I-5. Proposed new and modified ammonia nitrogen removal processes, South Lake Tahoe: Breakpoint chlorination (new).

will be removed by downstream breakpoint chlorination. The quantity of ammonia to be removed by breakpoint chlorination will vary from 5 to 16 mg/l, depending on the plant flow and temperatures.

Another approach to overcoming the limitations of the stripping process has been developed by CH2M/HILL Consulting Engineers.[8] Although the process is only in its initial stages of development, preliminary tests indicate it may be a significant advance in the state of the art of nitrogen removal. It appears that the new process overcomes most of the foregoing limitations and has the advantage of recovery of ammonia as a byproduct.

The improved process, shown diagrammatically in figure I-6, includes an ammonia-stripping unit and an ammonia-absorption unit. Both of these units are essentially sealed from the outside air but are connected by appropriate ducting. The stripping gas, which initially is air, is maintained in a closed cycle. The stripping unit operates essentially in the same manner that is now being or has been used in a number of systems, except that this system recycles the gas stream rather than using single-pass outside air.

Most of the ammonia discharged to the gas stream from the stripping unit is removed in the absorption unit. The absorbing liquid is maintained at a low pH to convert absorbed and dissolved ammonia gas to ammonium ion. This technique effectively traps the ammonia and also has the effect of maintaining the full driving force for absorbing the ammonia, since dissolved ammonia

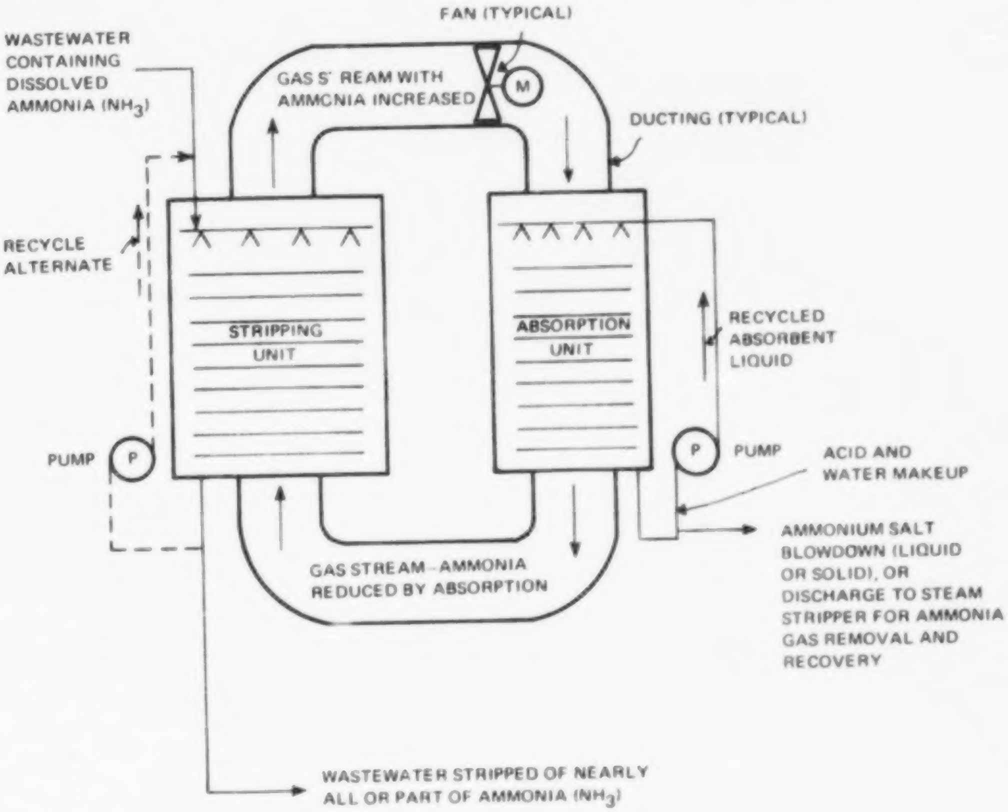
FAN (TYPICAL)
WASTEWATER CONTAINING DISSOLVED AMMONIA (NH_3)
GAS STREAM WITH AMMONIA INCREASED
M
DUCTING (TYPICAL)
RECYCLE ALTERNATE
STRIPPING UNIT
ABSORPTION UNIT
RECYCLED ABSORBENT LIQUID
PUMP
P
P
PUMP
ACID AND WATER MAKEUP
AMMONIUM SALT BLOWDOWN (LIQUID OR SOLID), OR DISCHARGE TO STEAM STRIPPER FOR AMMONIA GAS REMOVAL AND RECOVERY
GAS STREAM - AMMONIA REDUCED BY ABSORPTION
WASTEWATER STRIPPED OF NEARLY ALL OR PART OF AMMONIA (NH_3)

Figure I-6. Process for ammonia removal and recovery.

gas does not build up in the absorbent liquid. The absorption unit can be a slat tower, packed tower, or sprays similar to the stripping unit, but will usually be smaller owing to kinetics of the absorption process.

The absorbent liquid initially is water with acid added to obtain low pH, usually below 7.0. In the simplest case, as ammonia gas is dissolved in the absorbent and converted to ammonium ions, acid is added to maintain the desired pH. If sulfuric acid is added, for example, an ammonium sulfate salt solution is formed. This salt solution continues to build up in concentration and the ammonia is finally discharged from the absorption device as a liquid or solid (precipitate) blowdown of the absorbent. With current shortages of ammonia-based fertilizers, a salable byproduct may result.

Other methods of removal of the ammonia from the absorbent may also be applicable, depending on the acid used and the desired byproduct. Ammonia gas or aqua ammonia could be produced, for example, by steam stripping the absorbent. In this case, acid makeup would be unnecessary.

It is believed that the usual scaling problem associated with ammonia-stripping towers will be eliminated by the improved process, since the carbon dioxide which normally reacts with the calcium and hydroxide ions in the water to form the calcium carbonate scale is eliminated from the stripping air during the first few passes. The freezing problem is eliminated owing to the exclusion of nearly all outside air. The treatment system will normally operate at the temperature of the wastewater.

Chapter II

SELECTIVE ION EXCHANGE

The selective ion exchange process derives its name from the use of zeolites that are selective for ammonia relative to calcium, magnesium, and sodium. The zeolite currently favored for this use is clinoptilolite, which occurs naturally in several extensive deposits in the Western United States. Studies of the process have been conducted by Battelle Northwest[9] and the University of California.[10] Clinoptilolite used in studies conducted by Battelle Northwest for EPA was obtained from the Hector, Calif., leases of the Baroid Division of the National Lead Company, Houston, Tex. The clinoptilolite is crushed and sieved to obtain a 20 by 50 mesh size. Ammonia is removed by passing the wastewater through a bed of clinoptilolite at a rate of about 10 bed volumes per hour.

The use of clinoptilolite was investigated at the University of California with the objective of optimizing its application to ammonia removal from wastewaters. Pilot-plant operations were carried out at three different municipal sewage-treatment plants. An average ammonia removal of 96 percent was obtained in these operations with influent ammonia nitrogen concentrations of about 20 mg/l.

The ammonia capacity of the clinoptilolite was found to be nearly constant over the pH range of 4.0 to 8.0, but diminished rapidly outside this range. The effect of wastewater composition on the ammonia exchange capacity was analyzed by exhausting clinoptilolite beds with waters having different chemical compositions. For relatively constant influent ammonia concentrations, the ammonia exchange capacity was observed to decrease sharply with increasing competing action concentrations up to about 0.01 molar. Increases of cation concentrations above this value continued to decrease the exchange capacity, but to a much lesser degree. Ammonia removal to residual levels less than 0.5 mg/l ammonia nitrogen is technically feasible, but only with shorter service cycles and greater regeneration requirements. Flow rates in the range of 7.5 to 15 bed volumes per hour had no effect on ammonia effluent values.

Battelle Northwest conducted pilot studies of the clinoptilolite process applied to secondary effluents, advanced waste treatment effluents, and clarified raw sewage.[9,11] Ammonia removals ranging from 93 to 97 percent were demonstrated using a 100,000-gal/d mobile pilot plant. These studies were conducted at several different locations across the United States.

After about 150-200 bed volumes of normal-strength municipal waste have passed through the bed, the capacity of the clinoptilolite has been used to the point that ammonia begins to leak through the bed. At this point, the clinoptilolite must be regenerated so that its capacity to remove ammonia is restored.

The key to the applicability of this process is the method of handling the spent regenerant. The resin is regenerated by passing concentrated salt solutions through the exchange bed when the ammonia concentration has reached the maximum desirable level. Following regeneration, the ammonia-laden spent-regenerant volume is about 2.5 to 5 percent of the throughput treated before regeneration.

The original approach to recovering and reusing the regenerant was to use a lime slurry as the regenerant so that the ammonium stripped from the bed during regeneration would be converted to gaseous ammonia, which could then be removed from the regenerant by air stripping.[9]

Regeneration with lime alone was found to be a rather slow process; therefore, the ionic strength of the regenerant solution was increased by the addition of salt (NaCl). The increased ionic strength of the regenerant plus the presence of sodium ion accelerates the removal of ammonia from the zeolite. Although most of the sodium chloride added to the regenerant is converted to calcium chloride by continuous recycle of the regenerant, sufficient sodium ion remains under steady state conditions to promote the elution of the ammonium ions. The sodium ion has a higher diffusion coefficient than calcium ion, which is believed responsible for increasing the ammonia elution rate. With the lime-slurry regenerant, the regenerant stripping tower handles only a small fraction of the total plant throughput. Heating the stripping tower, even during cold weather periods, is then practical.

The use of the high-pH regenerant is accompanied by an operational problem. Some plugging of the bed with $Mg(OH)_2$ and $CaCO_3$ occurs when the high-pH regenerant is used. Attrition of the zeolite is aggravated by the violent backwashing needed to remove these solids, and is 0.17-0.25 percent per cycle, making makeup clinoptilolite costs a significant factor. These problems make more recently developed methods of regenerant recovery more attractive.

In one approach, ammonia in the regenerant solution may be converted to nitrogen gas by reaction with chlorine which is generated electrolytically from the chlorides already present in the regenerant solution. This process can be carried out with a regenerant of neutral pH so that the problem of precipitation of $Mg(OH)_2$ and $CaCO_3$ within the bed during regeneration is eliminated. Also, cold weather does not affect the regenerant recovery process. The regenerant solutions used are rich in NaCl and $CaCl_2$ which provide the chlorine produced at the anode of the electrolysis cell. The reactions for the destruction of ammonia by chlorine are the same as for breakpoint chlorination.

During regeneration of the ion exchange bed, a large amount of calcium is eluted from the zeolite along with the ammonia. This calcium may be removed from the spent regenerant solution by soda ash softening before passing the spent regenerant through the electrolytic cells. The softening step would lower the calcium concentration below the level that would cause calcium hydroxide formation in the electrolytic cells. High flow velocities through the electrolysis cells are required in addition to a low concentration of $MgCl_2$ to minimize scaling of the cathode by calcium hydroxide and calcium carbonate. Acid flushing of the cells would be necessary to remove this scale when the cell resistance becomes too high for economical operation.

In pilot tests of the electrolytic treatment of the regenerant at Blue Plains, Battelle Northwest found that about 50 Wh of power were required to destroy 1 gram of ammonia nitrogen (NH_3-N). When related to the treatment of water containing 25 mg/l NH_3-N, the energy consumed would be 4.7 kWh per 1,000 gallons. Tests at South Tahoe also indicated that a value of 50 Wh per gram is reasonable for design.[12] Preliminary capital and operating costs of $1.5 million and 9 cents per 1,000 gallons, respectively, were estimated by Battelle for a 10-mgd plant using electrolytic destruction of ammonia in recycled regenerant containing chloride salts of calcium, sodium, and magnesium. Electrolytic treatment of the regenerant avoids the disposal of ammonia to the atmosphere or disposal of aqueous ammonia concentrates. Total costs, including capital amortization, were estimated at 12.7 cents per 1,000 gallons.[11]

A 22.5-mgd plant designed by CH2M/HILL for the Upper Occoquan Sewage Authority in the State of Virginia will employ selective ion exchange with electrolytic treatment of the regenerant for ammonia removal. This plant will utilize soda ash softening of the regenerant to avoid cathodic scaling of the electrolysis cells. A simplified flow schematic of the regeneration system is illustrated in figure II-1. The regeneration of the clinoptilolite beds will be accomplished with a 2-percent solution of NaCl. The spent regenerant will be collected in a large holding tank to minimize variation in the calcium content before soda ash addition for calcium removal. After the soda ash addition, the regenerant will be clarified and transferred to another holding tank where the regenerant will be recirculated through electrolysis cells for ammonia destruction.

Design criteria for the ammonia-removal plant for the Upper Occoquan District are summarized in table II-1. The electrolysis cell to be used by this plant is a 500-Ampere unit manufactured by Pacific Engineering and Production Company of Nevada, Henderson, Nev. The cell consists of a lead dioxide coated graphite anode in a cylindrical stainless steel vessel which is the cathode. The lead dioxide is highly resistant to attack by chlorine or oxychloroacids. The estimated total cost for this plant is 12.6 cents per 1,000 gallons for the selective ion exchange process.

In order to develop the design criteria for the Occoquan plant, CH2M/HILL conducted pilot tests of the process at the South Tahoe plant.[12] The ammonia concentration in the wastewater at South Tahoe ranged from 21 to 28 mg/l during these pilot tests. After about 6 weeks of pilot-plant operation, the calcium concentration of the influent increased from about 55 mg/l to about 80 mg/l. This increased calcium concentration together with concurrently occurring lower influent temperatures reduced the quantity of ammonia that could be loaded onto the clinoptilolite before a breakthrough of 1 mg/l of ammonia. The average loading to the clinoptilolite column before breakthrough of 1 mg/l of ammonia was 144 bed volumes with an influent containing 55 mg/l calcium at 22° C. When the influent calcium increased to 80 mg/l and the temperature dropped to 14° C, the loading capacity of the clinoptilolite column dropped to 104 bed volumes. Ammonia removals achieved were in excess of 95 percent.

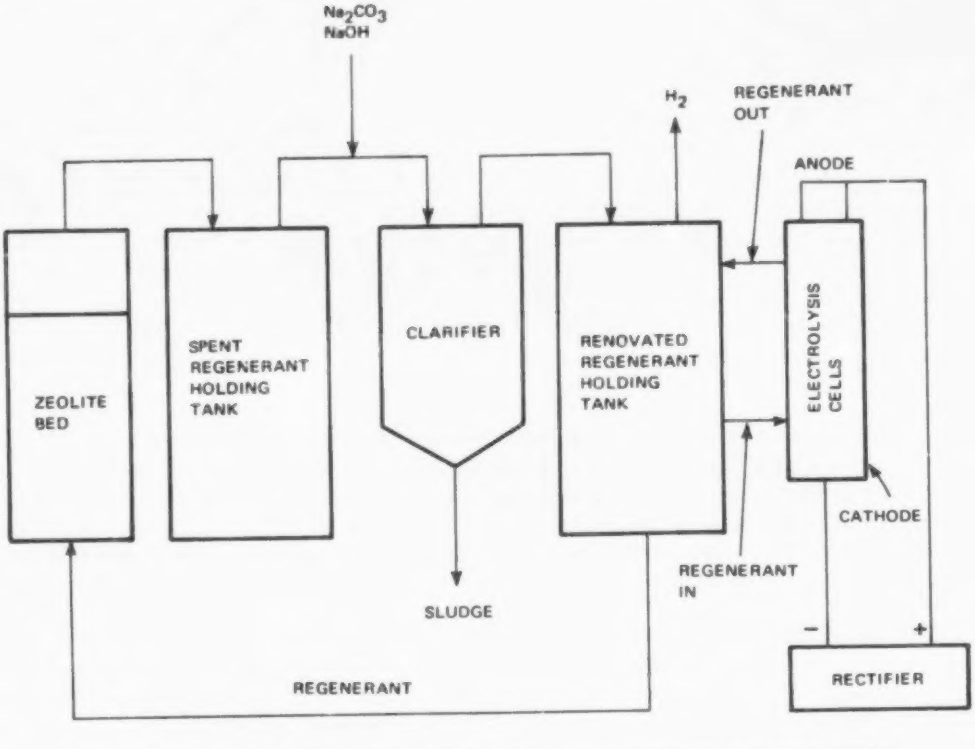

Figure II-1. Simplified flow diagram of Upper Occoquan regenerant treatment system.

Table II-1.—*Design criteria for the Upper Occoquan ammonia removal plant at 22.5-mgd flow rate*

Exchange beds:	
Size and type	10-foot-diameter × 50-foot-long horizontal pressure units
Number	8
Media	Clinoptilolite
Media size	20 × 50 mesh
Bed depth	4 feet
Bed length	50 feet
Bed width	10 feet
Service cycle loading:	
Average	9.1 BV/h
Maximum	14.1 BV/h
Hydraulic loading:	
Average	4.4 gal/min/ft^2
Maximum	6.9 gal/min/ft^2
Flow:	
Average	3.2 mgd per bed
Maximum	5 mgd per bed
Length of service cycle	200 BV
Bed loading	365 pounds NH_3 per bed cycle
Backwash water	Carbon-column effluent
Backwash rate	8 gal/min/ft^2
Exchange-bed regeneration:	
Length of cycle	3.1 hours
Regeneration rate	10 BV/h
Regenerant recovery:	
Method	Electrolysis
Power requirement	40 Wh per gram NH_3-N destroyed
NH_3 destruction rate	0.16 pound NH_3-N per hour per cell
Number of electrolytic cells in service	480
Total number of cells provided	720
Rectifiers:	
Number	3
Capacity	750 kW
Salt requirements	13,900 lb/d

The pilot column was regenerated successfully with a 2-percent sodium chloride solution at neutral pH. No loss of clinoptilolite by attrition was observed when using the neutral regenerant, and no difficulties in backwashing were observed. Although the neutral regeneration scheme was found to involve 30-40 bed volumes of regenerant rather than the 10 or less needed by others with the high-pH schemes, the minimization of attrition losses is achieved without significant disadvantage. The closed-loop regenerant-recovery system results only in added downtime for regeneration.

Scaling within the electrolytic cell used for regenerant recovery was the primary concern of the Occoquan pilot-plant study; therefore, the electrolytic cell was routinely dismantled and inspected for scaling. The flow rate through the cell was set initially at velocities of 0.13 to 0.16 ft/s, and a thin buildup of scale was observed on the cathode at the bottom-cell-inlet end after 160 hours of operation. After 230 hours of operation, the flow velocity was reduced to 0.06 ft/s, and very light scale buildup was observed depositing over the entire cathode area.

Scale was removed from a 1-in.^{2} area of the cathode, and the flow velocity through the cell was increased to 0.21 ft/s to determine the effect of scaling at higher cell velocities. At this increased flow, which was maintained for most of the period of the pilot-plant study, no new scale was deposited on the cathode. Visually, it appeared that from 25 to 50 percent of the previously deposited scale was removed. These observations suggest that scaling within the cell can be controlled by sufficient flow velocities. The average power requirements for regenerant recovery were measured as 43.3 Wh per gram ammonia destroyed. To allow for normal system losses, a design value of 50 Wh/g appears reasonable.

An alternative to air stripping or electrolysis of the regenerant is steam stripping. A 0.6-mgd plant in Rosemount, Minn., which is now entering its startup period, utilizes this technique.[13,14] At Rosemount ammonia is recovered from the spent ion exchange regenerant in an ammonia stripper. Steam is injected into a distillation column countercurrent with the regenerant solution to strip off the ammonia. An air-cooled plate-and-tube condenser then condenses the vapor for collection in a covered tank as 1-percent aqueous ammonia for sale as a fertilizer, However, it is a dilute (1 percent) ammonia solution, which reduces its potential for sale as a fertilizer, since commercial fertilizers require handling of only 1/10 the volume of liquid for the same ammonia application.

No detailed data on the Rosemount design and anticipated operating parameters were available at the time of this report. An EPA evaluation of the plant will be made in 1974 after the initial shakedown problems are resolved. The steam-stripping process is based on the use of the high-pH regenerant, which has the disadvantages noted earlier. Battelle Northwest's evaluation of steam stripping[a] indicates that it is economically feasible if the regenerant volume is held to 4 bed volumes per cycle, which is achievable with high-pH regenerant. The steam requirements were estimated to be 15 pounds per 1,000 gallons. At a steam cost of $2 per 1,000 pounds, the steam costs would be only 0.03 cent per 1,000 gallons. Heat recovery by contacting the cold regenerant with stripped regenerant and by contacting it with the condenser would be necessary to achieve economical operation. Because of the unstable, high-pH regenerant, scaling problems on the heat exchanges could be anticipated.

Another technique for regenerant recovery is the use of the stripping-recovery process (shown in fig. I-4) on the spent regenerant. A 6-mgd plant at North Lake Tahoe is being designed using this approach. Tests to date indicate that ammonia sulfate concentrations of 50 percent are readily achievable in the absorption tower. The estimated costs of the selective ion exchange approach based on this technique of regenerant recovery are shown in figure II-2. No credit for potential sale of ammonium sulfate has been included.

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[a] B. W. Mercer, Battelle Northwest, personal communication, Dec. 14, 1973.

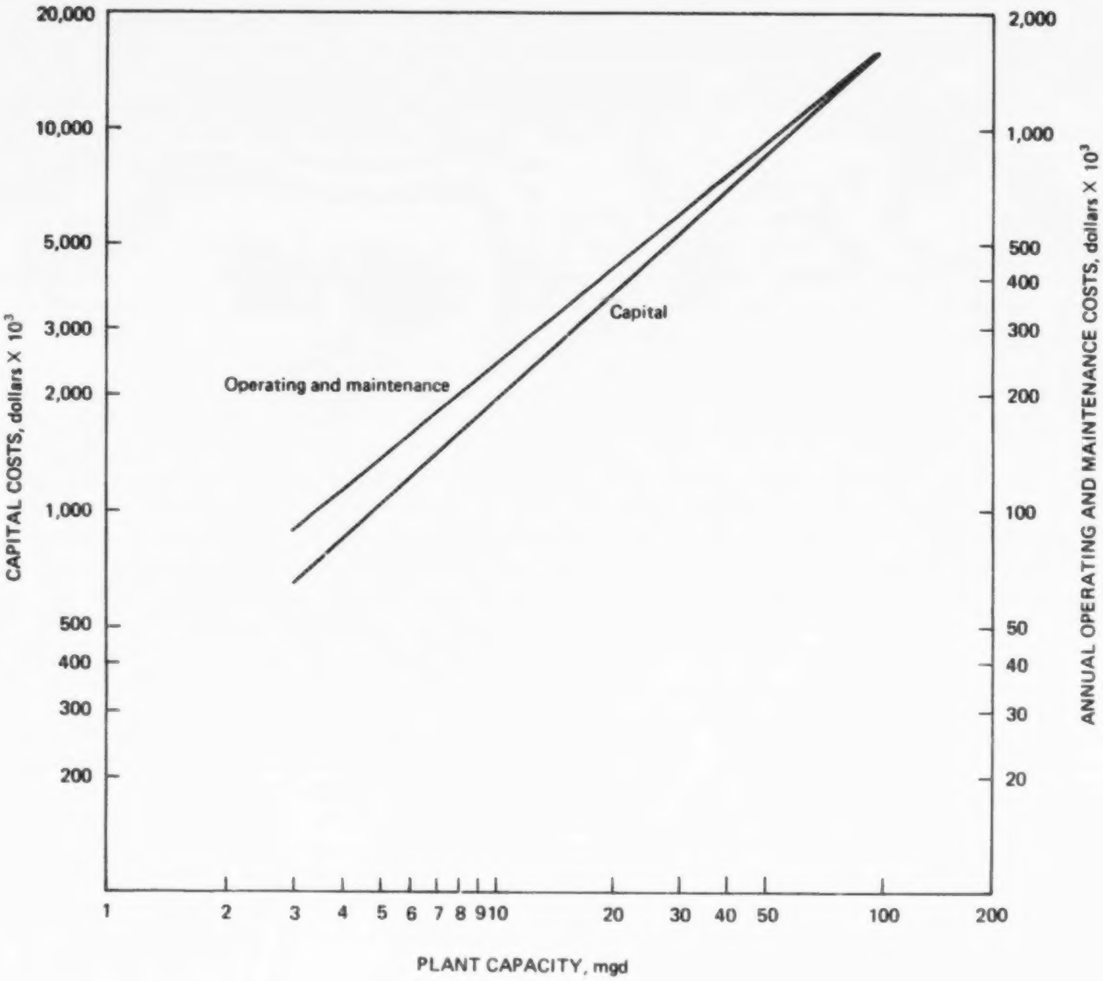

Figure II-2. Ammonia removal by selective ion exchange. (EPA STP Index = 200; includes engineering, legal, administrative, construction financing, and contingencies.)

Chapter III

BREAKPOINT CHLORINATION

When chlorine is added to a wastewater containing ammonia nitrogen, ammonia reacts with the hypochlorous acid formed to produce chloramines. Further addition of chlorine to the breakpoint converts the chloramines to nitrogen gas. The chlorine and ammonia reactions in dilute solutions are

$$NH_4 + HOCl \rightarrow NH_2Cl \text{ (monochloramine)} + H_2O + H^+$$

$$NH_2Cl + HOCl \rightarrow NHCl_2 \text{ (diochloramine)} + H_2O$$

$$NCHl_2 + HOCl \rightarrow NCl_3 \text{ (nitrogen trichloride)} + H_2O$$

The reactions are dependent on pH, temperature, contact time, and initial chlorine-to-ammonia ratio. Chlorine is added to the wastewater being treated until the chlorine residual has reached a minimum (the breakpoint) and the ammonia is removed. A typical breakpoint curve is shown in figure III-1. The reaction with ammonia is very rapid. Less than 1 minute, in the pH range of 7.0 to 8.0, and all of the free chlorine is converted to monochloramine at a 5:1 weight ratio of chlorine:ammonia nitrogen. As the weight ratio exceeds 5:1, the monochloramine breaks down and forms dichloramine and ammonia,

$$2NH_2Cl \rightarrow NHCl_2 + NH_3$$

Monochloramine is then oxidized by excess chlorine under slightly alkaline conditions to nitrogen gas,

$$2NH_2Cl + HOCl \rightarrow N_2\uparrow + 3HCl + H_2O$$

Stoichiometrically, a weight ratio of 7.6:1 of chlorine to ammonia nitrogen is required to oxidize ammonia to nitrogen gas.

Breakpoint chlorination tests on domestic wastewaters at the Blue Plains plant indicate that 95 to 99 percent of the ammonia is converted to nitrogen gas and that no significant amount of nitrous oxide is formed.[15] The quantity of chlorine required to achieve breakpoint was found to decrease with an increasing degree of treatment before the breakpoint process. The quantity of chlorine required for breakpoint chlorination of raw wastewater was found to be 10 parts by weight of Cl_2 to 1 part of NH_3 nitrogen. This ratio decreased to 9:1 Cl_2:NH_3 nitrogen for secondary effluents, and 8:1 Cl_2:NH_3 nitrogen for lime-clarified and filtered secondary effluent. The Blue Plains tests found that the chlorine dose was minimized at pH values between 6.0 and 7.0. The minimum NO_3 production (1.5 percent of the NH_3-N) occurred at pH 5.0. At pH 8.0, the nitrate production increased to 10 percent of the influent NH_3 nitrogen. NCl_3 production at the breakpoint decreased from 1.5 percent to the influent at pH 5.0 to 0.25 percent at pH 8.0. Temperature did not affect the product distribution or the required chlorine dose in the range 5° to 40° C.

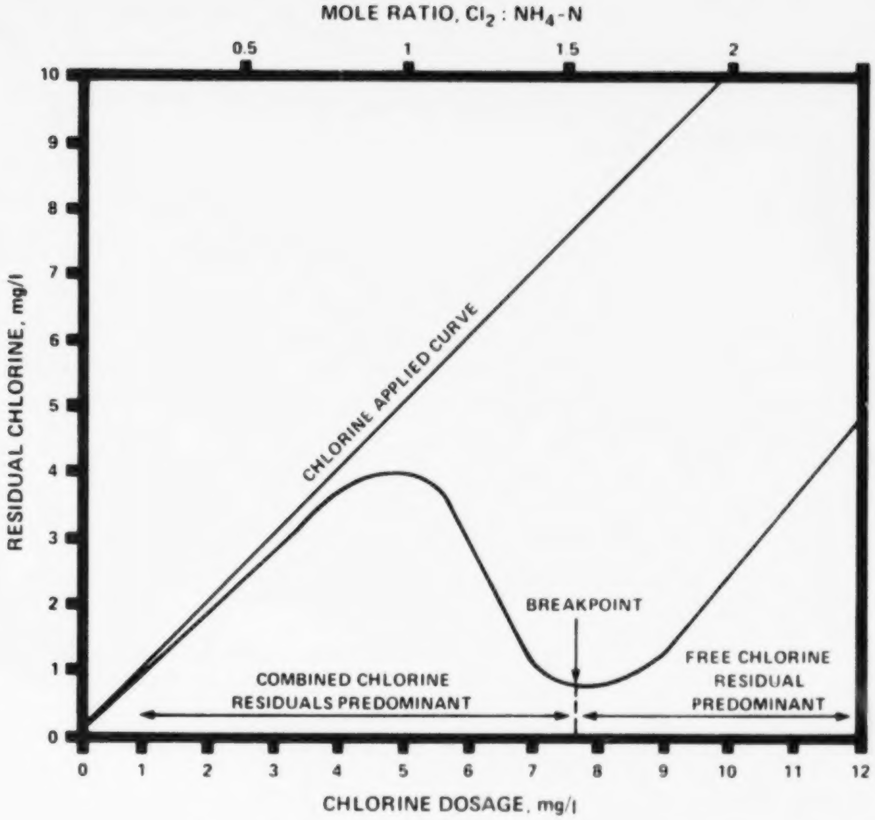

Figure III-1. Typical breakpoint-chlorination curve.

The use of chlorine produces an equivalent weight of hydrochloric acid which may depress the pH of the wastewater unless the natural alkalinity is adequate or a base such as sodium hydroxide is added. If the pH is allowed to fall, highly odorous nitrogen trichloride (NCl_3) is formed, which is an intolerable end product. If a base is used to prevent pH depression, the mixing of the wastewater, chlorine, and base must be extremely violent to avoid local areas of low pH which would generate NCl_3. Tests at Blue Plains showed that eductors do not give adequate chlorine-wastewater mixing, which did result in localized low-pH regions in which objectionable quantities of NCl_3 formed. Violent mechanical mixing is required. The use of sodium hypochlorite rather than chlorine does not depress the pH and avoids the foregoing problem.

The use of chlorine gas may produce more acid than can be neutralized by the wastewater. According to the EPA study reported by Pressley,[15] 14.3 mg/l of alkalinity (as $CaCO_3$) are required to neutralize the acid produced by the oxidation of 1 mg/l NH_3-N to N_2. Either sodium hydroxide or lime may be used for pH control if the wastewater is deficient in alkalinity. A wastewater containing 25 mg/l NH_3-N requires an alkalinity of about 357 mg/l if chlorine gas is used.

A significant factor in considering this process for application in some cases is the addition of dissolved solids inherent to the process. If, for example, chlorine gas were used and the influent ammonia nitrogen concentration were 25 mg/l, the dissolved solids would be increased by 156 mg/l. Neutralizing with lime would result in a total increase of 306 mg/l of total solids. If the chlorinating agent were sodium hypochlorite, the increase in dissolved solids would be 177 mg/l.[16]

The effects of breakpoint chlorination on organic nitrogen are somewhat uncertain. The Blue Plains tests[15] found only a "slight reduction in organic nitrogen within the two hour contact time." Other tests[17] observed a decrease in organic nitrogen content as the Cl_2:N ratio increased. Reductions from 3.2-3.5 mg/l to 0.2-0.4 mg/l organic nitrogen were reported for the breakpoint process. The authors,[17] however, felt that such apparent removals result from an analytical anomaly in which the organochloramine formed is not measured as nitrogen in the Kjeldahl organic nitrogen analysis. At higher chlorine dosages, however, their literature review indicated that organochloramines will be oxidized to aldehydes and nitrogen gas. The breakpoint reactions of organochloramines proceed more slowly than the ammonia chloramines, and probably will not be complete in a 30-minute contact time.

Several recent studies[16,17,18,19] have investigated the possibility of adding only enough chlorine to form monochloramines and then removing the monochloramines on activated carbon. Some advantages would be realized if monochloramine could be removed by activated carbon. The theoretical Cl:N ratio for 100 percent ammonia removal would drop from 7.6:1 for breakpoint to about 5:1 for the formation of monochloramine. The dissolved solids added to the system and the alkalinity requirements would be significantly reduced. Two studies[16,17] found that ammonia removals of about 50 percent could be achieved at Cl:N ratio of 5:1 when the breakpoint process was followed by activated-carbon adsorption. Complete removal still required dosages of about 9:1 in three studies.[16,17,18] Carbon contact times of 10 minutes were found to be adequate for complete dechlorination of the effluent.[16]

Experiences with the breakpoint process in South Africa[20] confirm that automatic control of the process is important. The African researchers concluded that monitoring of the ammonia coupled with automatically controlled chlorine dosing is a necessity. A successful, automated-computer-control system has been developed and demonstrated at the Blue Plains pilot plant.[21] This system matches the quantity of chlorine fed to the quantity of incoming nitrogen, and also controls the pH to 7.0 to minimize the formation of NCl_3 and NO_3. (See fig. III-2.)

There are several projects in the design or construction stage utilizing the breakpoint-chlorination process. The 7.5-mgd South Lake Tahoe plant is adding facilities to provide breakpoint chlorination of the quantities of ammonia which escape the upstream nitrogen-removal processes (5-16 mg/l).[6] The Orange County, Calif., 15-mgd wastewater reclamation plant now nearing completion will include facilities to remove the 2-3 mg/l of ammonia that will escape the upstream ammonia-stripping process.[22] Chlorine gas will be supplied from purchased 1-ton cylinders and by an on-site electrolytic generator rated at 2,000 lb/d. The chlorine generation system will utilize an electrochemical cell to electrolyze sodium chloride brine to chlorine gas and sodium hydroxide solution. The sodium hydroxide solution will be used in an adjacent sea water desalting plant.

A 60-mgd facility is under design for Montgomery County, Md., by CH2M/HILL, which will utilize the breakpoint process as the primary nitrogen-removal process. In this plant, sodium hypochlorite will be produced on site by electrolysis of a salt brine. The Cortland, N.Y., 10-mgd physical-chemical plant design includes facilities for breakpoint chlorination of the portion of the flow required to meet stream standards.

The costs of the process applied to the 309-mgd plant at Blue Plains were estimated at 6.7 cents per 1,000 gallons, with chemical costs constituting 5.9 cents of this value. These costs were based on a chlorine cost of only $75 per ton and a dose of only 120 mg/l. The control of pH was assumed to be by lime addition (1 pound of lime per pound of chlorine) at a lime cost of $24 per ton. In any case, the cost of the chlorine itself constitutes a large portion of the total project costs. Assuming a chlorine cost of 0.07 cent per pound and a Cl:N ratio of 8:1, the chlorine cost for removal of 25 mg/l ammonia would be 11.8 cents per 1,000 gallons. The chlorine demand for this dose is equivalent to 1,668 lb/mg.

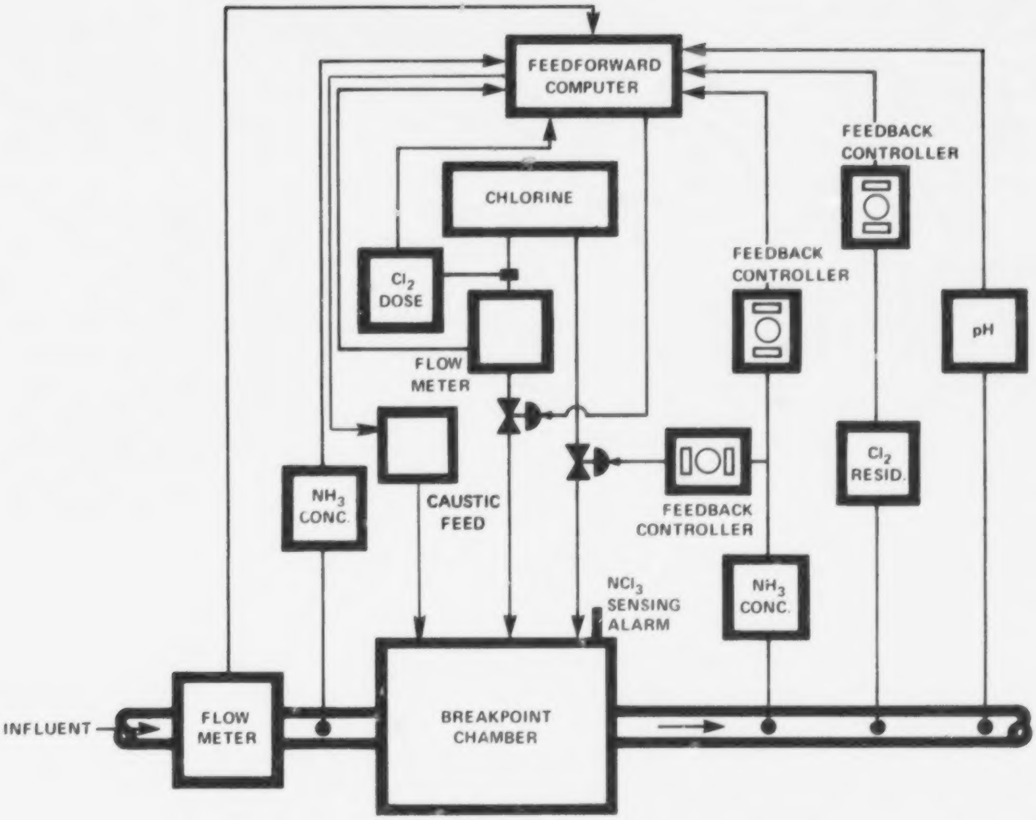

Figure III-2. Breakpoint-chlorination control system.

The breakpoint process is useful for eliminating low concentrations of ammonia as a polishing step following another nitrogen-removal process.

Chapter IV

COMPARISON OF PROCESSES

Each of the processes discussed earlier has its advantages and disadvantages. Unfortunately, no single process for nitrogen removal is superior to others both in terms of performance and economics.

The ammonia-stripping process has the advantages of low cost, removal of ammonia with a minimal addition of dissolved solids, simplicity, and reliability. However, it has the disadvantages of poor efficiency in cold weather and the potential for scaling problems that may reduce its efficiency, and it raises concerns, whether valid or not, over ammonia gas discharge. The new stripping-recovery system overcomes many of these problems, but at the sacrifice of low process costs.

The selective ion exchange process has the advantages of high efficiency, insensitivity to temperature fluctuations, removal of ammonia with a minimal addition of dissolved solids, and the ability to eliminate any discharges of nitrogen to the atmosphere other than nitrogen gas. This process has the disadvantage of relatively high cost, and process control and operation are relatively complex.

The breakpoint chlorination process has the advantages of low capital cost, a high degree of efficiency and reliability, insensitivity to cold weather, and the release of nitrogen as nitrogen gas. It has the disadvantage of adding a substantial quantity of dissolved solids to the effluent in the process of removing the ammonia, it will raise public concerns over handling of chlorine gas, the process controls required are relatively complex, and it requires a downstream dechlorination process.

The relative costs of the physical-chemical nitrogen processes for a 10-mgd plant are

- Ammonia stripping, 5 cents per 1,000 gallons
- Selective ion exchange, 10-13 cents per 1,000 gallons
- Breakpoint chlorination, 11 cents per 1,000 gallons

These costs all are based on the removal of 25 mg/l ammonia nitrogen. The cost of biological nitrogen removal by the three-stage activated-sludge process has been estimated[23,24] at about 13 cents per 1,000 gallons. Preliminary estimates on the costs of the new ammonia-stripping/ammonia-recovery process discussed earlier, which minimizes the seasonal restrictions on the ammonia-stripping process, indicate that the cost will be 8-10 cents per 1,000 gallons. It can be seen from the above costs that there is little economic incentive to select one process over another if faced with a requirement for cold weather removal of ammonia. The choice must be made by weighing the advantages and disadvantages of each approach in light of the circumstances applicable to a specific project.

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SLUDGES GENERATED IN PHOSPHATE REMOVAL PROCESSES

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TOKYO, JAPAN

FEBRUARY 1974

SLUDGES GENERATED IN PHOSPHATE REMOVAL PROCESSES*

Aluminum and iron salts and lime have been used in the physical-chemical treatment of wastewater to remove phosphate. Use of these chemicals increases the mass of sludge and affects its dewatering properties.

Quantity

Isgård,(1) in discussing Swedish experience, has presented a table which relates sludge production to chemical dose (Table 1). It is believed that he was referring to experience with tertiary treatment. The quantities of alum and iron sludges are easily calculated from the chemical equations showing the reactions of Al^{3+} and Fe^{3+} (see Table 2). The dose of Al or Fe used is related to the phosphorus level in the wastewater. It usually ranges from 1.8 to 2.2 atoms per atom of P. Isgård's figures are seen to be quite reasonable.

When aluminum and iron salts are added at the primary clarification stage, much more primary sludge is produced than predicted by the chemical equations. The reason is the substantial increase in the efficiency of the clarifier. Tables 3 and 4 show the anticipated increase in sludge mass when Al and Fe are added at various points in the wastewater treatment process.

The illustration in Table 3 compares sludge production when Al and Fe are added to the primary clarifier with sludge production when they are added to the aerator. Less sludge is produced when the chemicals are added to the aerator. This is true because the BOD load on the aerator is much reduced when the chemicals are added to the primary. Since the BOD load is reduced, there is much less conversion of organic material to carbon dioxide; consequently total mass of sludge is higher.

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* Presented at the Third U.S./Japan Conference on Sewage Treatment Technology, Tokyo, Japan, February 14, 1974.

The mass of sludge formed when lime is added to wastewater can also be calculated from the chemical equations (see Table 5). It is necessary to know the chemical analysis of the water before treatment and estimate it after treatment. The major difference in calculation method is that the initial dose of $Ca(OH)_2$ is estimated from the alkalinity of the wastewater rather than from a relationship between the phosphorus level in the wastewater and the $Ca(OH)_2$ dose. Figure 2 shows a correlation of lime dose with alkalinity of the wastewater.

Dewatering Properties

There are a few generalizations that can be made about the sludges generated by Al, Fe, and $Ca(OH)_2$ addition. Aluminum and iron salts generally reduce the solids content of the primary clarifier sludge. Sludges cannot be thickened to as high a solids content. The effect is often unnoticed at lower doses, but when the ratio of metal dose to wastewater suspended solids increases, the effect is substantial. For example, if enough Al or Fe is added to a typical United States wastewater to remove 90 percent of the phosphate, the sludge thickening and dewatering properties will be noticeably poorer.

With lime, the effect is the opposite. Lime improves thickening and dewatering properties substantially. It is often possible to eliminate sludge conditioning agents. Mass of sludge is generally much higher but volume is often lower than if Al or Fe salts were used.

Table 6 attempts to express these qualitative statements in a semi-quantitative way. It should only be used as an approximate guideline for anticipated results.

Examples

Barrie. Tests were reported by Ian Gray(2) of the Ontario Ministry of the Environment (Canada). The plant conditions are as follows:

3.0 MGD (11,400 m^3/da)
bar screens, grit removal
primary clarifier
activated sludge
final clarifier
two stage digestion
waste activated sludge recycled to plant inlet

Alum addition to the primary was compared to alum addition to the aerator.

A high dose of alum (200 mg/l of alum, equivalent to 200 x 0.091 = 18.2 mg/l Al) significantly lowered sludge solids.

If alum dose was less than 150 mg/l of alum, there was no effect on sludge density.

The point of alum addition (to the primary or to the aerator) did not affect sludge density.

Little River. This is a small primary plant (15,000 m^3/da) in the city of Windsor, Ontario.(3) Data are summarized in Table 7. Alum and lime addition are compared to each other and to a control period (the previous year). Alum addition caused a slight reduction in filter yield and cost of conditioning chemicals increased. Lime caused a substantial increase in sludge concentration and filter yield, and reduced chemical conditioning costs.

Blue Plains. This is the principal plant in Washington, D. C. Table 8 describes the processing sequence at this plant. During a lengthy test period, 40 mg/l of alum (9.1% Al) was added to the aerators. Effects are summarized below:

A. Thickening. Sludge is 7.5%. No change.

B. Digestion. Sludge leaves at 3.5%. No change. No interference with digestion.

C. Elutriation. Higher polymer dose is needed to retain fines. Cost is $7/dry ton vs. $3/dry ton.

D. Filtration. Filter yield is about the same, but solids content of cake is 20% vs. 23%. Chemical cost is $7/dry ton, slightly over $1/dry ton higher than previously.

Contra Costa, California. This system utilizes chemical treatment in the primary clarifier for phosphorus removal, followed first by carbonaceous removal and nitrification in a single aeration stage, and finally by denitrification.(4) Lime will be used to pH 11 with 14 mg/l ferric chloride. The lime addition reduces the biological load to the second stage, allowing longer sludge age needed for nitrification. It also provides heavy metals removal, and offsets the acid formed in nitrification.

Waste biological solids are returned to the primary clarifier. Solids in the primary clarifier underflow are 5-9 percent. Sludge is first centrifuged in a solid bowl centrifuge without polymer. The cake contains most of the calcium carbonate, and the centrate contains most of the calcium phosphate, magnesium hydroxide, and organic material. The centrate is once again centrifuged using a higher rotational speed and polymer. The first cake can be calcined to produce reusable lime. The cake from the centrate is incinerated for disposal.

Salt Lake City. Addition of Al, Fe, and $Ca(OH)_2$ have been investigated on a large pilot plant scale by the Eimco Corporation at Salt Lake City, Utah.(5) Results are summarized in Tables 9 and 10. Results for the Al and Fe cases are based on a very limited amount of data.

Results indicate excellent sludge characteristics with lime. Results with Al and Fe are poor, primarily because the experimenters could not thicken the sludge. This work has continued. Much more information has been collected but is not yet available. A modification in the design of the clarifier has allowed the Al and Fe sludges to settle to higher solids, giving better filter yields.

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(5) Burns, D. E., and G. L. Shell, "Physical-Chemical Treatment of a Municipal Wastewater Using Powdered Carbon," Environmental Protection Technology Series, EPA-R2-73-264, Aug. 1973.

TABLE 1 -- SLUDGE PRODUCED
BY CHEMICAL ADDITION

ALUM : ADD 4 x mg/l of Al

FERRIC: ADD 2.5 x mg/l of Fe

LIME : ADD 1 to 1.5 mg/l of $Ca(OH)_2$

(AFTER E. ISGÅRD - SWEDISH EXPERIENCE)

TABLE 2 -- Al and Fe SLUDGE PRODUCTION

$Al + PO_4 = AlPO_4$
1 kg 4.52 kg

$Al + 3OH = Al(OH)_3$
1 kg 2.89 kg

$Fe + PO_4 = FePO_4$
1 kg 2.70 kg

$Fe + 3OH = Fe(OH)_3$
1 kg 1.91 kg

Table 3 : Calculated Sludge Mass (lb/M.G.)

	Conventional	Fe to Primary	Fe to Aerator	Al to Aerator	Al to TF Clarifier
Primary					
SS Removal	50%	75%	50%	50%	50%
Sludge Solids	1250	1875	1250	1250	1250
Fe Solids	0	605			
Al Solids	0				
Total	1250	2480	1250	1250	1250
Activated Sludge					
Secondary Solids	715	536	804	804	
Fe Solids			541		
Al Solids				425	
Trickling Filter					
Secondary Solids	(656)				745
Al Solids					483
Totals	1965	3016	2595	2479	2478

Table 4 Basis for Sludge Mass Calculation in Table 3

Cation/P Dose (mol/mol)	lb Chemical Sludge/lb Cation lb/lb Al	 lb/lb Fe
1.5	3.9	2.4
1.75	3.8	2.3

Assumptions:
Cation/P Dose = 1.5 mol/mol to aerator
Cation/P Dose = 1.75 mol/mol to primary or before Trickling Filter clarifier

Influent Sewage
BOD = 230 mg/l
SS = 300 mg/l
P = 10 mg/l

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TABLE 5 -- REACTION OF LIME WITH WASTEWATER

$$5Ca(OH)_2 + 3PO_4 = Ca_5OH(PO_4)_3 + 9OH$$

$$Ca(OH)_2 + HCO_3 = CaCO_3 + H_2O + OH$$

$$Mg + 2OH = Mg(OH)_2$$

$$Ca(OH)_2 = Ca + 2OH$$

TABLE 6 -- FILTRABILITY OF PHOSPHATE SLUDGES

CHEMICAL	PRIMARY		PRIMARY + WAS	
	YIELD	COST*	YIELD	COST*
Al^{3+}	0.5-0.8	MUCH MORE	0.7-1.0	MORE
Fe^{3+}	0.5-0.8	MUCH MORE	0.7-1.0	MORE
LIME	1.1-1.5	LESS	1.3-1.8	MUCH LESS

* COST OF CONDITIONING CHEMICAL

TABLE 7 -- LITTLE RIVER WPCP, WINDSOR, PRIMARY PLANT (15,000 m^3/da)

	CONTROL	ALUM	LIME
DOSE (mg/1)	0	150	125-150
SLUDGE (mg/1)	158	310	338
SLUDGE CONCN.(%)	6.2	5.8	11.1
FILTER YIELD (kg/m^2-hr)	25	23	35
CONDITIONING COST* ($/metric ton)	17.70	19.60	13.00

* $FeCl_3$ and LIME

TABLE 8 -- BLUE PLAINS WPCP, WASHINGTON, D.C.

CAPACITY	300 MGD (1,130 x $10^3 m^3$/da.)
PROCESS	PRIMARY CLARIFICATION HIGH-RATE ACTIVATED SLUDGE

SLUDGE PROCESSING

A. PRIMARY IS MIXED WITH WASTE A.S. and THICKENED
B. DIGESTION
C. TWO-STAGE ELUTRIATION
D. FILTRATION (POLYMER + $FeCl_3$)

TABLE 9 -- SALT LAKE CITY PILOT PLANT, PRIMARY ONLY

	$FeCl_3$	ALUM	$Ca(OH)_2$
DOSE (mg/l)	120	150	460
SLUDGE (mg/l)	168	144	840
SLUDGE CONCN.(%)	1.3	0.3	12.0
GRAVITY THICKENING			
SOLIDS LOADING (kg/m^2-da)	74	24	210
UNDERFLOW SOLIDS (%)	3.0	1.5	20.0

TABLE 10 -- SALT LAKE CITY PILOT PLANT, PRIMARY ONLY

	$FeCl_3$	ALUM	$Ca(OH)_2$
DOSE (mg/l)	120	150	460
VACUUM FILTRATION			
$Ca(OH)_2$ DOSE (kg/kg)	0.2	0.2	0
YIELD (kg/m^2-hr)	5.9	3.9	49
CAKE SOLIDS (%)	20	20	40

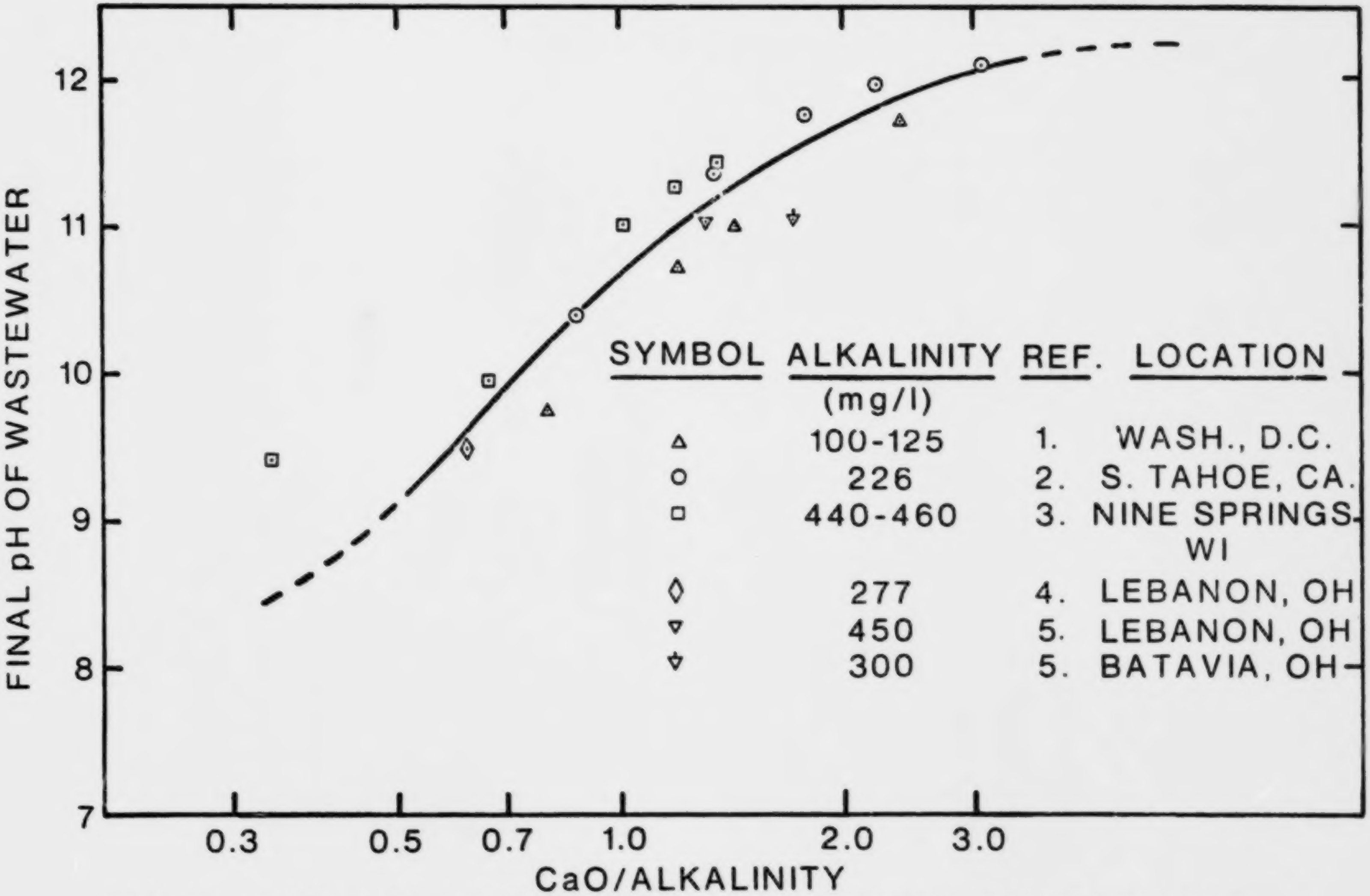

FIG. 2: RATIO OF LIME DOSE (mg/l) TO INITIAL WASTEWATER ALKALINITY (mg/l)

EPA EXPERIENCES IN OXYGEN-ACTIVATED SLUDGE

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THIRD U.S./JAPAN CONFERENCE ON SEWAGE TREATMENT TECHNOLOGY
TOKYO, JAPAN

FEBRUARY 1974

*The presentation was based on material prepared by Richard C. Brenner for the U.S. Environmental Protection Agency - Technology Transfer Design Seminar Program

EPA EXPERIENCES IN OXYGEN-ACTIVATED SLUDGE

Richard C. Brenner

INTRODUCTION

Utilization of oxygen aeration for activated sludge treatment is receiving increasing attention in wastewater treatment plant construction in the United States. The concept, although more than 20 years old, has received serious consideration only during the last six years with the development of several cost-effective systems for dissolving and utilizing oxygen gas in an aeration tank environment.

The rapid transition from the drawing boards to full-scale implementation has been possible because of intensive government and private research and development programs. The U.S. Environmental Protection Agency (EPA) and its predecessor organizations have contributed significantly to the total research and development effort. The purpose of this paper is to summarize the role of EPA during the period of 1968-1974 as the oxygen aeration process progressed to its current level of development.

As outlined in Table 1, EPA has pursued seven active projects to date. The projects include in-house pilot plant studies to examine process kinetics, extramural feasibility grants and contracts, extramural materials and safety projects, and extramural demonstration grants.

The EPA contribution to the projects described in Table 1 exceeded $3.4 million through Fiscal Year 1974 (ended June 30, 1974). The cost breakdown by project is given in Table 2.

Test facilities, experimental plans, and results (where available) for each of the above projects are summarized in the following sections.

THE BATAVIA PROJECTS

A research and development contract was awarded to the Union Carbide Corporation in October 1968 to evaluate a proprietary staged, covered-tank oxygenation system at the Batavia, New York, Water Pollution Control Plant. Union Carbide was awarded a follow-up contract in June 1970 to

TABLE 1. EPA RESEARCH AND DEVELOPMENT PROJECTS ON OXYGEN AERATION

	Project	Objective
1.	Batavia I and II (Union Carbide Corporation)	Establish feasibility of multi-stage, covered-tank oxygenation concept.
2.	Newtown Creek (New York City)	Scaled-up demonstration of multi-stage, covered-tank oxygenation system.
3.	Las Virgenes (California) Municipal Water District	Demonstration of single-stage, covered-tank oxygenation system.
4.	FMC Corporation	Establish feasibility of open-tank oxygenation concept.
5.	EPA/District of Columbia (Blue Plains) Pilot Plant	Determine process kinetics over wide range of operating conditions.
6.	Bureau of Reclamation	Materials of construction corrosion testing.
7.	Rocketdyne Division of Rockwell International	Define safety requirements and develop safety manual and checklist.

TABLE 2. EPA RESEARCH AND DEVELOPMENT EXPENDITURES ON OXYGEN AERATION THROUGH FY-73

Project	Cost to EPA	Type of Project
Union Carbide Corporation (Batavia I and II)	$ 795,000	Contracts
New York City (Newtown Creek)	$1,574,000	Grant
Las Virgenes (California) Municipal Water District	$ 186,000	Grant
FMC Corporation	$ 142,000	Grant
EPA/District of Columbia (Blue Plains) Pilot Plant	$ 500,000	Contracts and Inhouse
Bureau of Reclamation	$ 165,000	Contract
Rocketdyne Division of Rockwell International	$ 92,000	Contract
TOTAL	$3,454,000	

better define soluble organic removals and excess biological sludge production and to undertake initial pilot plant studies on oxygen sludge dewatering and stabilization. The oxygenation system was installed in one of two existing air-activated sludge trains at Batavia. During the first contract, the performance of the oxygen train was evaluated against that of the intact air train. A schematic diagram of the Batavia Plant after installation of the oxygen system is shown in Figure 1.

The oxygen system configuration evaluated at Batavia was the first large-scale embodiment of the now well known "UNOX" process.* A typical three-stage "UNOX" aerator is shown schematically in Figure 2. The aerator operates as a series of completely mixed stages, thereby approximating plug flow. Oxygen gas is fed under the aeration tank cover at the inlet end of the tank only and flows co-currently with the liquid stream from stage to stage. Gas is recirculated in each stage by centrifugal compressors which force the gas down hollow shafts out through submerged rotating spargers. Submerged turbines maintain suspension of the mixed liquor solids and disperse the oxygen gas. A mixture of unused oxygen gas, cell respiration by-product carbon dioxide, and inert gases is exhausted from the final stage, typically at an oxygen composition of about 50% and a flow rate equal to 10-20% of the incoming gas flow rate. Using co-current gas and liquid flow to match the decreasing dissolution driving force inherent in continually decreasing oxygen gas composition with the decreasing oxygen demand of wastewater undergoing biological treatment has proven to be a very efficient oxygen contacting and utilization technique.

A second-generation multi-stage process has been developed and utilized both by Union Carbide and Air Products and Chemicals, Inc. This adaptation of the original covered-tank concept replaces the recirculating compressors and rotating spargers with surface aerators. Oxygen transfer is accomplished by gas entrainment and dissolution. Submerged turbines are also used optionally where tank geometry requires additional mixing capability. As shown in Figure 3, all other aspects of the system are unchanged. The first Air Products and Chemicals version of the covered-tank, surface-aerator concept has been operating for approximately three years at the Westgate Treatment Plant in Fairfax County, Virginia (design flow 12 mgd). Operation commenced in July 1972

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*Mention of a trade name or commercial products does not constitute Environmental Protection Agency endorsement or recommendation for use.

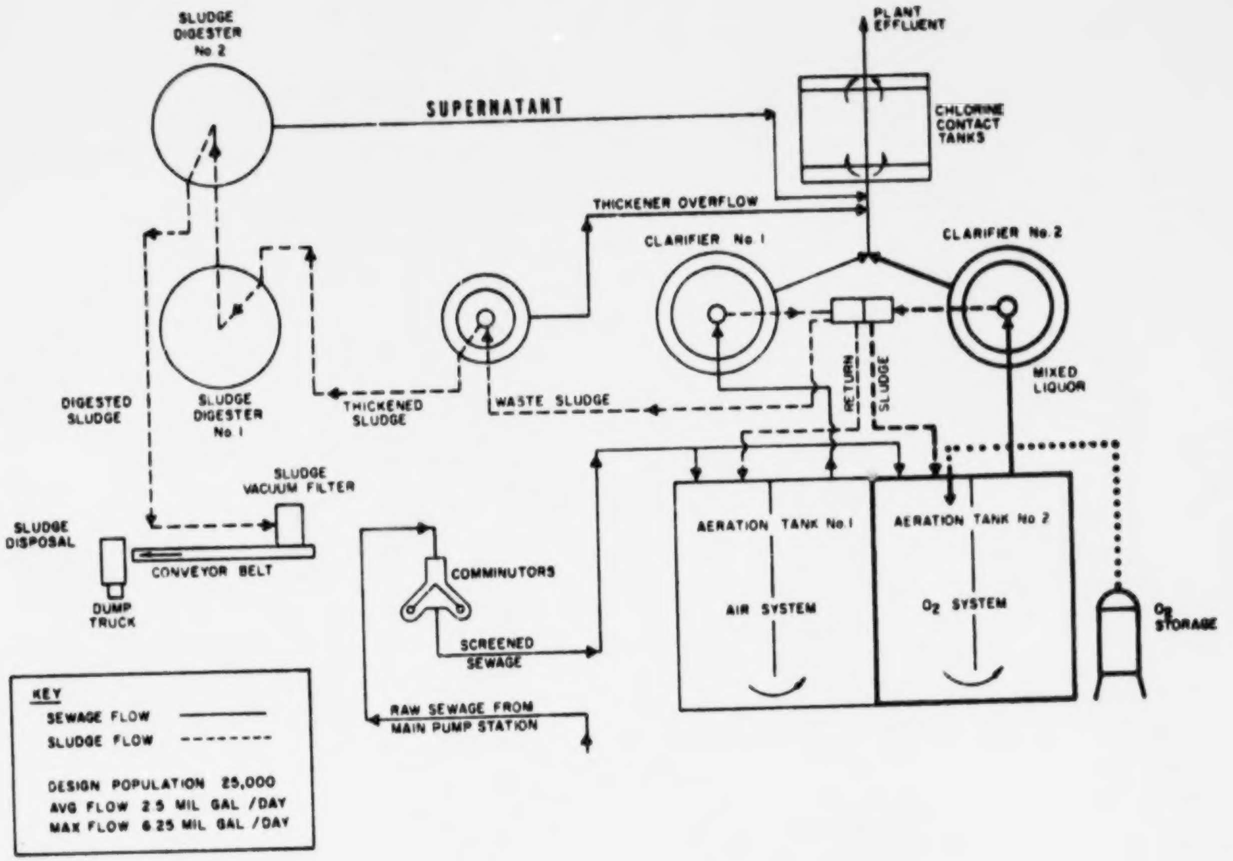

FIGURE 1. SCHEMATIC FLOW DIAGRAM FOR WATER POLLUTION CONTROL PLANT, CITY OF BATAVIA, NEW YORK

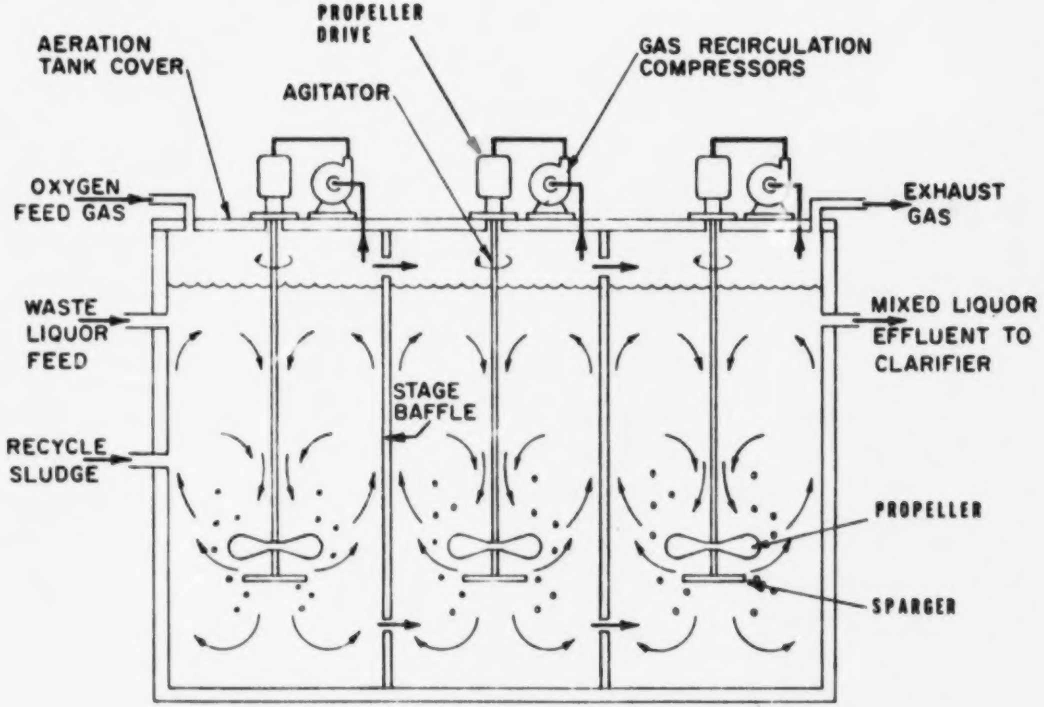

FIGURE 2. SCHEMATIC DIAGRAM OF MULTI-STAGE, COVERED-TANK OXYGENATION SYSTEM WITH GAS RECIRCULATION COMPRESSORS AND SUBMERGED TURBINE/SPARGERS

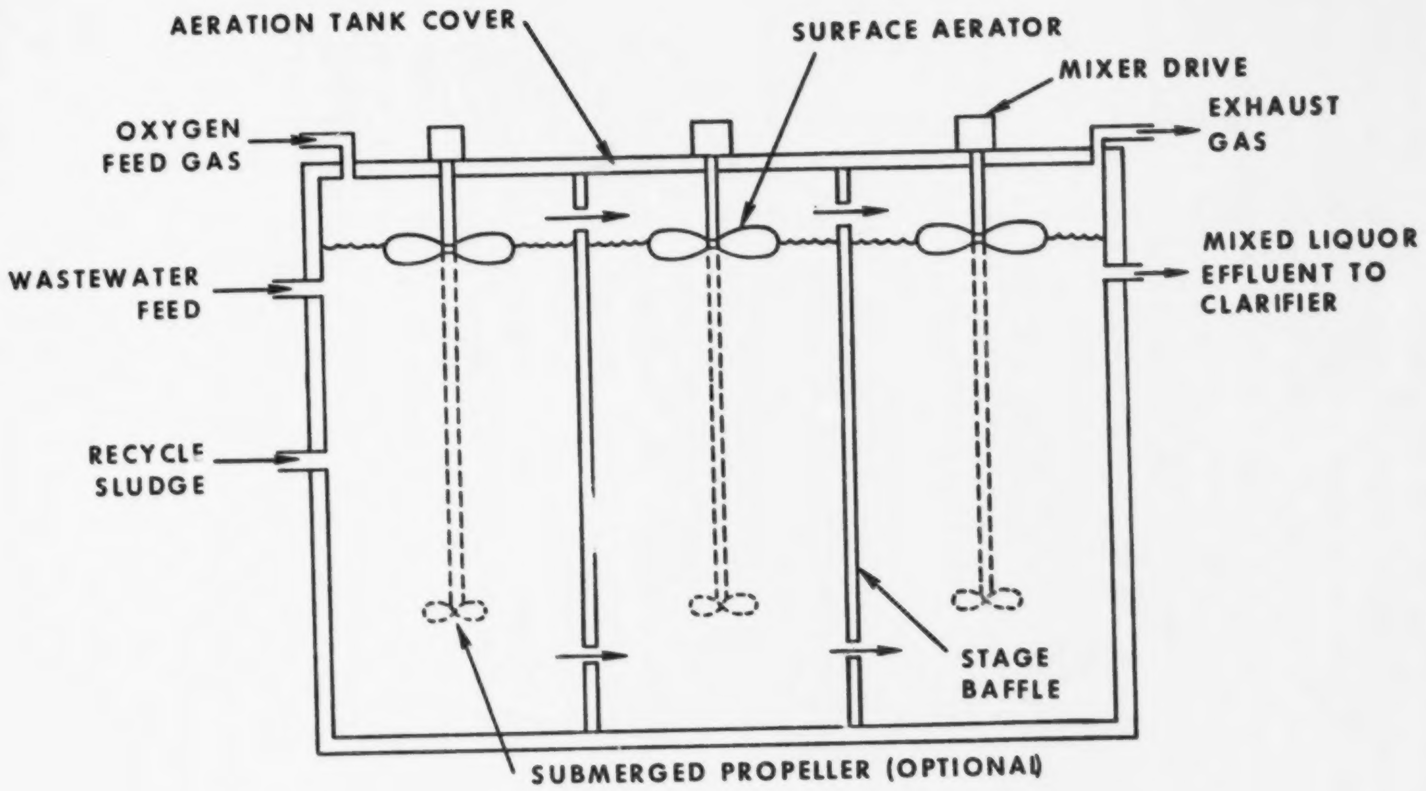

FIGURE 3. SCHEMATIC DIAGRAM OF MULTI-STAGE, COVERED-TANK OXYGENATION SYSTEM WITH SURFACE AERATORS

at Speedway, Indiana, the first municipal plant to utilize the Union Carbide surface aerator system (design flow 7.5 mgd). The surface aerator modification of the basic multi-stage process has exhibited better cost effectiveness for tanks up to approximately 15 feet deep and is being used increasingly in full-scale design. Market forecasts and actual experience to date of firms selling multi-stage, covered tank oxygen systems indicate that 80-85% of the plants that eventually utilize this oxygen system concept will employ surface aerator designs. A report currently being prepared for EPA by Air Products and Chemicals documenting the Fairfax County, Virginia case history from inception through two years of operation with the oxygen system will be available by mid-1975.

The results of the Batavia projects have been widely disseminated in two EPA Water Pollution Control Research Series Reports (17050 DNW 05/70) (17050 DNW 02/72). One of the conclusions expressed in these reports is that oxygen aeration can provide equal treatment efficiency to air aeration with only one-third as much aeration volume. This conclusion has been subject to widespread criticism. In that this generalization was reached by comparing an efficient oxygen contacting system with a relatively inefficient coarse-bubble air aeration system, the criticism appears to be justified. The increasing variety of air aeration equipment being marketed offers a wide range of oxygen transfer kinetics. Some of this equipment has been shown to be capable of supporting the higher mixed liquor solids concentrations necessary for justifying smaller volume biological reactors (e.g., the INKA aeration system, Divet, et al., 1963). Design engineers are urged to investigate and prepare cost estimates for both air and oxygen systems as a basis for process selection. Process selection should be made from a total integrated system comparison, including aeration, secondary clarification, and excess biological sludge handling and disposal requirements.

Pertinent results of the two Batavia projects relating only to oxygen system performance are summarized below:

1. The feasibility of achieving high oxygen gas utilization (91-95%) was established.
2. Efficient biological performance (90-95% BOD_5 and suspended solids removals, 80-85% COD removal) was demonstrated with short aerator detention periods (1.4-2.8 hours based on Q) and high organic volumetric loadings (140-230 lb BOD_5/day/1,000 ft^3).

3. High mixed liquor dissolved oxygen (D.O.) levels (8-12 mg/l) were maintained at high mixed liquor suspended solids (MLSS) concentrations (3,500-7000 mg/l).
4. Warm weather secondary clarifier performance deteriorated above an average surface loading of 1,600 gpd/ft^2.
5. Oxygen sludge exhibited excellent thickening properties during secondary clarification (settled sludge of 1.5-3.0% solids).
6. Aerobic digestion of oxygen waste activated sludge with oxygen produced comparable volatile suspended solids (VSS) reduction rates to those given in the literature for air aerobic digestion processes. Reductions in oxygen sludge VSS concentrations of 25 and 40% were achieved with 7 and 15 days of aerobic stabilization, respectively.
7. Direct vacuum filtration of undigested oxygen waste activated sludge was shown to be feasible using 10% ferric chloride for conditioning. Cake yields of 3.5-4.5 lb/hr/ft^2 and moisture contents of 83-85% were achieved at a cycle time of 2.4 min/rev. Moisture content improved to 75-80% but cake yield dropped to 1.5-2.5 lb/hr/ft^2 at a cycle time of 6.3 min/rev.
8. Vacuum filtration of aerobically digested oxygen waste activated sludge proved to be infeasible with the chemical conditioners tested.

THE NEWTOWN CREEK PROJECT

Results of the initial Batavia contract were judged sufficiently encouraging to justify a scaled-up demonstration of the multi-stage oxygen system in a large municipal plant. A research and development grant was awarded to New York City in June 1970 to convert one of sixteen parallel bays at its Newtown Creek facility to oxygen using the recirculating compressor/submerged turbine version of the "UNOX" process. The design flow of the test bay is 20 mgd, roughly 10 times higher than the capacity of the Batavia oxygen system. In addition to the $1.574 million EPA grant, New York City provided over $1.2 million in city funds in support of the project.

The Newtown Creek plant was designed on the modified aeration principle for 1.5 hours of aeration time (based on Q) and treatment efficiencies in the range of 65-70%. The city is now confronted with an upgrading problem in a land-locked neighborhood (see Figure 4), a situation common to many

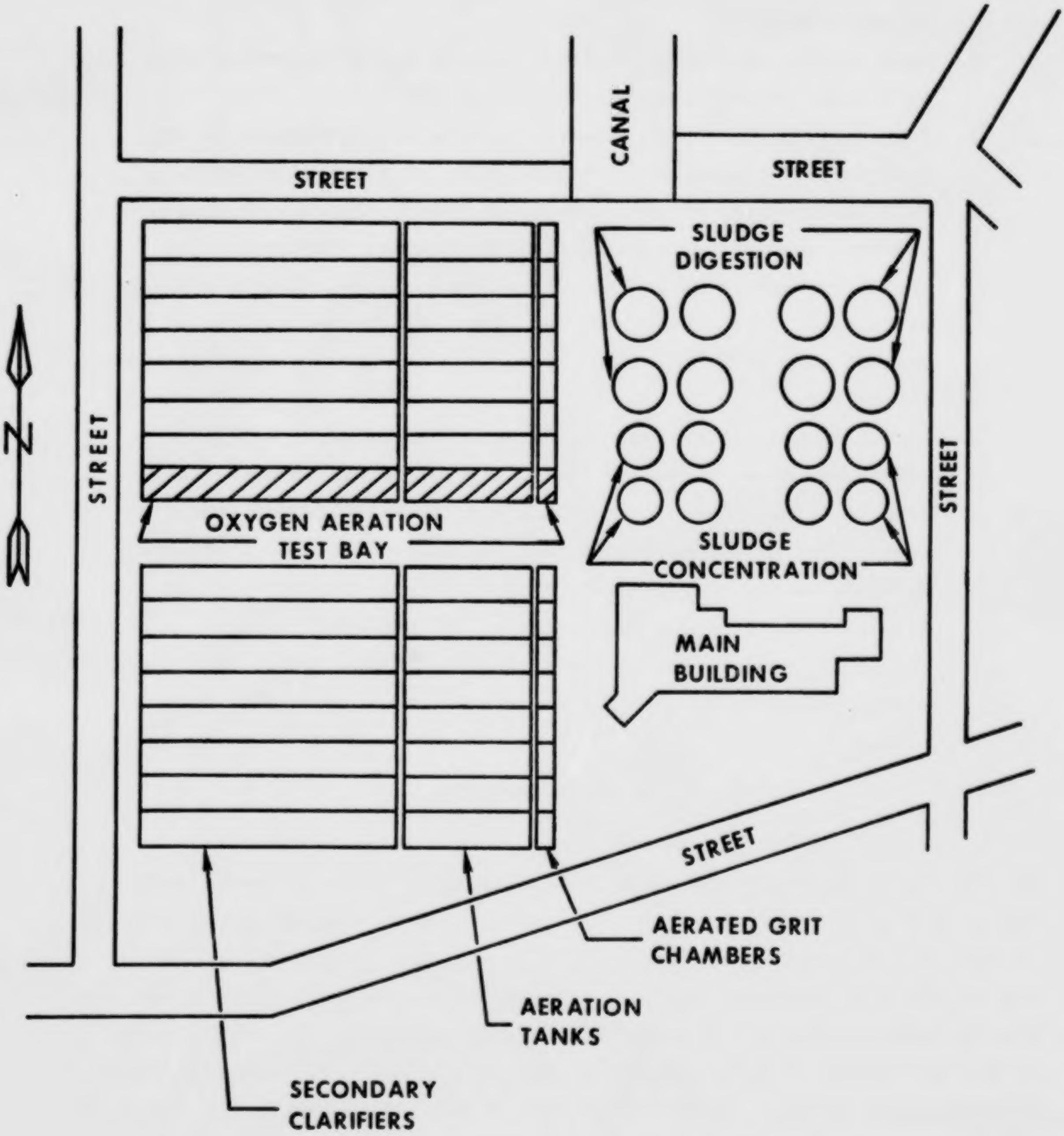

FIGURE 4. PLANT LAYOUT FOR NEWTOWN CREEK POLLUTION CONTROL PLANT, BROOKLYN, NEW YORK

large urban plants in the United States. Oxygen was believed to be a good candidate for achieving the required 90% BOD_5 removal within the confines of the existing aeration tanks and secondary clarifiers. Future conversion of the entire 310 mgd facility to oxygen was the ultimate objective provided a removal up to the 90% ± BOD_5 level could be consistently demonstrated in the test bay. Two views of the Newtown Creek test bay are shown in Figure 5.

The test bay went on stream in early June 1972. Extensive mechanical problems and unreliable meters prevented accurate data collection during the start-up phase (June 4, 1972-September 16, 1972) during which time the influent flow was increased from 11 mgd to the design level of 20 mgd. Data for this period are limited to effluent quality as summarized in Table 3.

TABLE 3. EFFLUENT QUALITY AT NEWTOWN CREEK
DURING START-UP (6/4/72-9/16/72)

		Effluent Concentration (mg/l)				
Flow (mgd)	Duration (weeks)	Total BOD_5	Soluble BOD_5	Total COD	Soluble COD	Suspended Solids
11 ± 2[a]	4	10	4	68	50	18
14 ± 3[b]	5	8	3	55	41	19
20 ± 4[c]	6	15	4	62	47	18

[a] MLSS = 4,865 mg/l, Detention Time (based on Q) = 2.7 hr ±.

[b] MLSS = 5,920 mg/l, Detention Time (based on Q) = 2.1 hr ±.

[c] MLSS = 4,260 mg/l, Detention Time (based on Q) = 1.5 hr ±.

Metering difficulties were finally resolved by mid-September 1972 permitting commencement of the extensive data collection program planned for this project. From September 17, 1972 through September 1, 1973, seven phases of a ten-phase experimental program were completed. The influent flow conditions for these seven phases are summarized in Table 4. Diurnal peak, average, and minimum flow rates for Phases 4 through 7 are given in Table 5. With the exception of Phase 7, the diurnal fluctuation patterns were selected to simulate the actual influent flow patterns of the Newtown Creek facility.

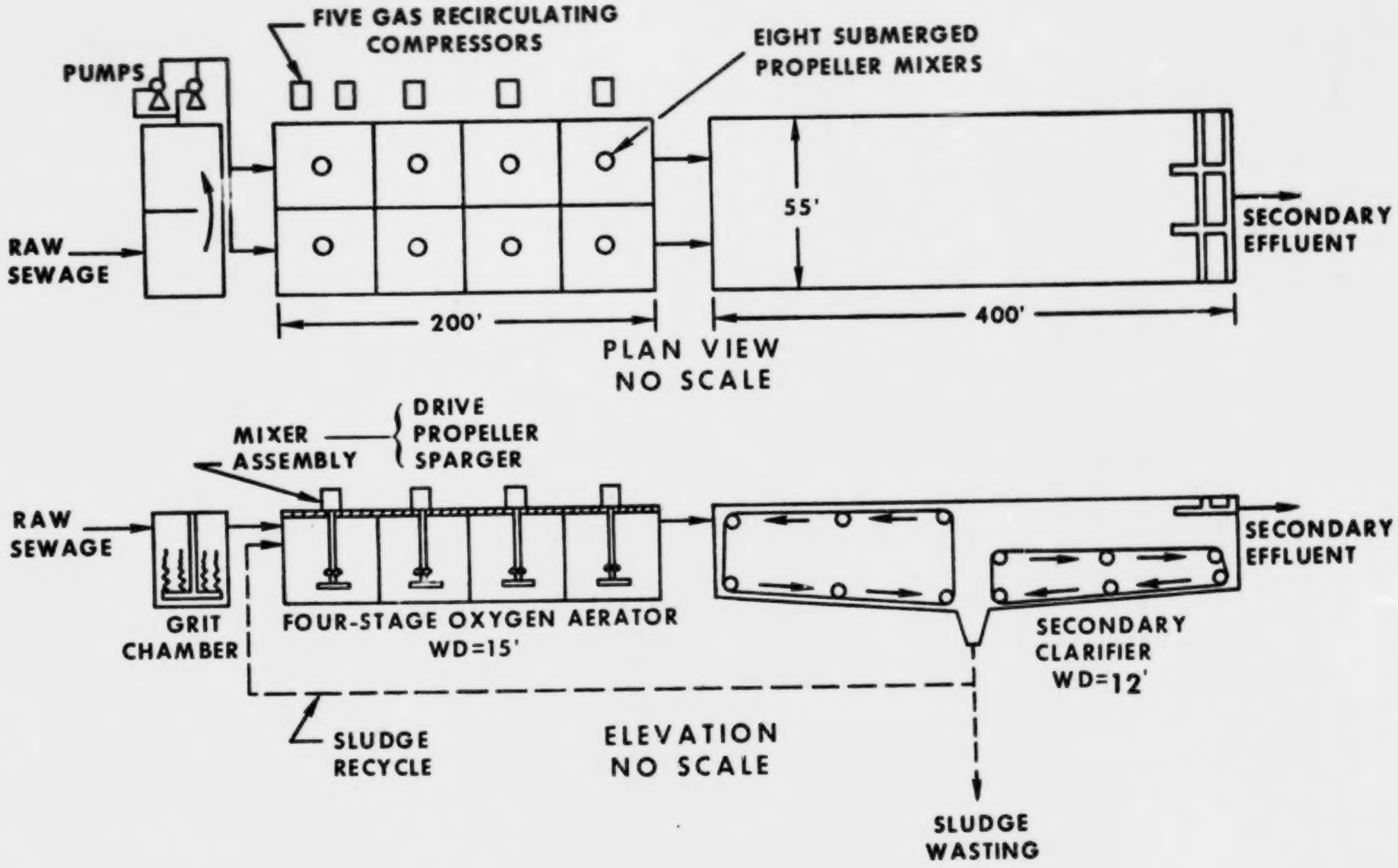

FIGURE 5. PLAN AND ELEVATION VIEWS OF OXYGEN AERATION TEST BAY, NEWTOWN CREEK

TABLE 4. EXPERIMENTAL SCHEDULE FOR NEWTOWN CREEK PROJECT (9/17/72-9/1/73)

Phase	Dates	Influent Flow Condition
1	9/17/72 - 11/25/72	20.8 mgd (Constant)
2	12/10/72 - 2/1/73	14 ⟶ 20 ⟶ 15 mgd Avg. 17.7 mgd (Winter Upset)
3	2/18/73 - 4/7/73	8 ⟶ 20 mgd AVG. 15.1 mgd (Winter Restart)
4	4/8/73 - 6/2/73	20.6 mgd (Diurnal)
5	6/3/73 - 7/7/73	25.3 mgd (Diurnal)
6	7/8/73 - 8/11/73	30.0 mgd (Diurnal)
7	8/12/73 - 9/1/73	35.4 mgd (Diurnal)

TABLE 5. DIURNAL FLUCTUATION PATTERNS FOR NEWTOWN CREEK (4/8/73-9/1/73)

Phase	Avg. Flow (mgd)	Peak Flow (mgd)	Minimum Flow (mgd)
4	20.6	24	14
5	25.3	30	17
6	30.0	36	19
7	35.4	37.5*	30

*Maximum influent pumping capacity.

A performance summary for the oxygenation system for Phases 1 through 7 is presented in Table 6. System sludge characteristics, aerator loadings, and secondary clarifier loadings are summarized in Tables 7, 8, and 9, respectively.

TABLE 6. PERFORMANCE SUMMARY FOR NEWTOWN CREEK
(9/17/72-9/1/73)

	Phase						
	1	2	3	4	5	6	7
Total BOD_5 In (mg/l)*	156	157	152	171	213	218	212
Total BOD_5 Out (mg/l)	9	21	17	17	22	21	23
% Removed	94	87	89	90	90	90	89
Soluble BOD_5 In (mg/l)*	84	78	91	102	113	99	88
Soluble BOD_5 Out (mg/l)	4	13	12	11	13	11	15
% Removed	95	83	87	89	88	89	83
Total COD In (mg/l)*	356	365	365	365	307	290	308
Total COD Out (mg/l)	61	88	76	77	70	64	62
% Removed	83	76	79	79	77	78	80
Soluble COD Out (mg/l)	50	69	63	69	58	49	46
Susp. Solids In (mg/l)*	149	146	144	159	147	125	131
Susp. Solids Out (mg/l)	12	22	17	18	24	17	17
% Removed	92	85	88	89	84	86	87
Sewage Temp. Range (°F)	77 ↓ 56	64 ↓ 53	51 ↓ 61	53 ↓ 69	62 ↓ 74	71 ↓ 77	72 ↓ 78

*No primary sedimentation. Concentrations shown are for raw sewage influent to oxygen aerator.

TABLE 7. AVERAGE SYSTEM SLUDGE CHARACTERISTICS FOR NEWTOWN CREEK (9/17/72-9/1/73)

Phase	MLSS (mg/l)	MLVSS (mg/l)	Return Sludge Flow (% of Q)	Return Sludge TSS (mg/l)	SVI (ml/gram)	SRT (days)
1	4,890	4,110	30	16,260	45	-*
2	5,060	4,150	40	12,835	59	-*
3	4,000	3,200	50	11,370	77	-*
4	3,875	3,110	45	13,420	53	1.35
5	4,550	3,640	44	15,975	42	1.37
6	4,155	3,340	34	16,330	43	1.24
7	3,090	2,485	25	12,685	48	0.77

*Sludge wasting data not reliable.

TABLE 8. AVERAGE AERATOR LOADINGS FOR NEWTOWN CREEK (9/17/72-9/1/73)

Phase	Detention Time -Based on Q- (hr)	F/M Loading $\left(\frac{\text{lb BOD}_5\text{/day}}{\text{lb MLVSS}}\right)$	Volumetric Organic Loading $\left(\frac{\text{lb BOD}_5\text{/day}}{1{,}000\ \text{ft}^3}\right)$
1	1.43	0.65	163
2	1.68	0.57	140
3	1.96	0.57	110
4	1.44	0.92	178
5	1.17	1.19	272
6	0.99	1.62	331
7	0.84	2.44	379

TABLE 9. AVERAGE SECONDARY CLARIFIER LOADINGS FOR NEWTOWN CREEK (9/17/72-9/1/73)

Phase	Surface Overflow Rate (gpd/ft^2)	Mass Loading $\left(\frac{\text{lb TSS/ft}^2}{\text{day}}\right)$	Weir Loading (gpd/ lineal ft)
1	945	50.1	129,000
2	805	48.2	110,000
3	686	32.6	93,000
4	936	43.7	127,000
5	1,150	63.0	157,000
6	1,364	63.3	186,000
7	1,609	52.0	219,000

From the beginning of the project, New York City officials considered performance of the oxygen test bay during cold weather the most critical segment of the experimental program. It was during the prolonged severe weather period that the true upgrading potential of oxygen for Newtown Creek would be most evident. A discussion of the data collected through September 1, 1973 at Newtown Creek is first prefaced, therefore, with a summary description of the operational difficulties encountered during the 1972-73 winter season.

During Phase 1 (autumn 1972), operation went smoothly and performance was obviously excellent. On November 25, 1972, the test bay was shut down temporarily to replace a bearing on the sludge recycle pump. What was planned to be a two-day outage turned into a two-week shutdown when the stocked spare bearing proved to be the wrong size and a new one had to be located. During the outage, sludge in the reactor was continually oxygenated while the sludge in the secondary clarifier was devoid of oxygen. Due to the imminence of the upcoming cold weather, it was decided to restart the system using the existing sludge rather than empty the tanks and take the time necessary to generate a new biomass. The oxygen test bay was put back into service on December 10, 1972.

For the first several weeks of Phase 2, performance was satisfactory as the influent flow rate was gradually increased to the design level of 20 mgd. Shortly thereafter sludge settling properties began to deteriorate and effluent BOD_5, COD, and suspended solids residuals exhibited a slowly increasing trend. Microscopic examination of the mixed liquor revealed the appearance of filamentous organisms of both apparent bacterial and fungus origin. Influent flow was then decreased in several increments during the month of January 1973 in an attempt to starve or "burn out" the filamentous culture or cultures and reestablish a "healthy" population.

Instead of eradicating the filamentous organisms, reduction of flow and organic loading seemed to have the opposite effect of stimulating proliferation. This proliferation was accompanied by the usual indicators of a bulking sludge, i.e., a substantial increase in SVI, a rising sludge blanket in the secondary clarifier, increasing suspended solids carry-over in the final effluent, and operational difficulty in managing total system sludge inventory.

In trying to determine the source of the filamentous intrusion, it was postulated that the bacterial species (Sphaerotilus) may have developed in the secondary clarifier sludge blanket during the aforementioned shutdown. A local pharmaceutical firm is known to discharge mycelia into the Newtown Creek sewer system and this was suspected as the source of the fungus organisms. By the end of January 1973, with the influent flow reduced to 15 mgd, the SVI had risen from a summer background level of 45-50 to 85-100, effluent suspended solids were exceeding 30 mg/1, effluent soluble BOD_5 had increased to over 20 mg/1, and the clarifier sludge blanket was continuing to rise. At this point a decision was made to "dump" the entire sludge inventory, hose all settled sludge pockets out of the reactor and clarifier, and start over. This second shutdown began on February 1, 1973.

After taking some additional time to make repairs to the sludge collection mechanism while the clarifier was dewatered, Phase 3 commenced on February 18, 1973. Conservative loading rates were utilized initially based on the premise that the best chance to prevent a reoccurrence of the filamentous condition was a program of gradual and modest increases in F/M loading until the 20 mgd design flow rate was reached. Relatively high MLSS concentrations of 5,000 mg/1 ±

were maintained as a further measure to minimize F/M loading. However, within several weeks a repeat of the experience encountered in Phase 2 became evident with the initial appearance of filaments in the oxygen sludge. This time the organisms were definitely identified as fungus. One possible reason for explaining the higher 1972-73 winter incidence and enrichment pattern of these fungus organisms in the Newtown Creek oxygen sludge as opposed to that facility's air sludges is the lower mixed liquor pH inherent to the operation of the covered-tank oxygen system.

With the prospect of impending project failure a real possibility, a joint decision (New York City, Union Carbide, and EPA) was reached to accept the presence and proliferation of the fungus organisms as a cold weather phenomenon and attempt to find an operating mode that would permit satisfactory winter performance at design flow in spite of them. Accordingly, a program of increased sludge wasting was initiated which eventually lowered the MLSS concentration to less than 4,000 mg/l and the SRT* to slightly more than one day. At the same time influent flow was elevated in several fairly rapid increments to 20 mgd (equivalent to an aerator detention time of 1.5 hours based on Q). These steps yielded an F/M at the end of the phase in the range of 0.75-0.80, considerably higher than the average of 0.57 for all of Phase 3. The altered operating philosophy proved to be the correct decision, resulting in a controllable clarifier sludge blanket and stable cold weather performance at design flow. Percentage removals for the remainder of Phase 3, although not as high as Phase 1, were within satisfactory limits. As the wastewater temperature increased during early spring, the concentration of filamentous organisms diminished, and they eventually disappeared in early May 1973.

During the summer of 1973, the Newtown Creek oxygen system exhibited remarkable capability for absorbing high hydraulic and organic loadings while still producing a high quality secondary effluent. Four diurnal phases (Phases 4-7) conducted from April 8, 1973 through September 1, 1973 successively increased the average influent flow from 20.6 to 35.4 mgd. During Phase 7 the average nominal aerator detention time was only 52 minutes with corresponding average F/M, volumetric, and clarifier surface loadings of

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* Defined as lb VSS under aeration/lb VSS wasted in the waste sludge and final effluent/day.

2.44 lb BOD_5/day/lb MLVSS, 379 lb BOD_5/day/1,000 ft^3, and 1,609 gpd/ft^2, respectively. These results confirmed and exceeded the high-rate loading potential of oxygen-activated sludge first seen at Batavia.

At this point in the project it was possible to offer the following interim status remarks:

1. The high-rate loading capability (nominal aeration time $\leq$ one hour) of oxygen aeration operating on Newtown Creek wastewater during warm weather was conclusively demonstrated.
2. Prospects appeared promising that a modified method of operation evaluated in late winter 1972-73 could circumvent the negative effects of what may be an indigenous cold weather filamentous condition with oxygen at Newtown Creek and permit satisfactory performance at a flow rate at least equal to the design level of 20 mgd (1.5 hours of nominal aeration time).
3. The operational measures employed to effect the improved performance in late winter 1972-73, namely high F/M's and low SRT's, occurred naturally to an even greater degree during the high loading phases of summer 1973.
4. The above comments provide a tentative basis for speculating that in some cases oxygen aeration may most beneficially be employed at ultra high loading rates substantially exceeding any which have been approved to date by State agencies.

Because of the importance attached to winter operation and performance, the project was extended to the end of April 1974. The two major questions which were to be addressed during the extended period were whether filamentous organisms (particularly fungus) would again infest the oxygen sludge as wastewater temperature dropped and, if so, would the modified method of winter operation previously described permit continuous efficient performance with a diurnal loading pattern centered around an average influent flow rate of 20 mgd. If the first few months progressed without upset, the flow rate was to be increased to 25 mgd and subsequently to 30 mgd in the last 2-3 months of the winter season. The reason for holding this latter option open was that if a year-round loading capability of 30 mgd could be demonstrated, the Newtown Creek Treatment Plant could conceivably be satisfactorily upgraded by converting only 11 or 12 of the existing 16 bays from air aeration to oxygen aeration.

Data for the extended operating period (September 1973 through April 1974) are summarized along with the first seven project phases in a paper entitled "Upgrading New York City Modified Aeration with Pure Oxygen." This paper was prepared by New York City personnel (Nash, et al.) and presented at the 47th Annual Conference of the Water Pollution Control Federation. It is recommended that both this Technology Transfer report and the New York City paper be reviewed in evaluating the Newtown Creek project. The project will also be extensively documented and analyzed in the final grant report now being prepared by the City. It is anticipated this report will be available for distribution by mid-1975.

Another aspect of the project is discussed briefly below. Initially the performance of the four-bed Pressure Swing Adsorption (PSA) Oxygen Generator was less than satisfactory. During the 1972 summer startup phase the unit was out of service due to mechanical problems roughly 40 percent of the time. These problems have since been largely corrected and the generator now functions with a down-time that varies between 5 and 10 percent. During the 1972 startup difficulties one of the four beds inadvertently became "loaded up" with water vapor. Subsequently, the maximum achievable output of the unit was 10 tons of gas per day at 90 percent oxygen purity versus a design output of 16.7 tons of gas per day at 90 percent oxygen purity. This necessitated an increase in consumption of and reliance on the back-up liquid oxygen reservoir during peak oxygen demand periods. The simplified three-bed (moving parts decreased 50 percent) PSA unit installed at Speedway, Indiana, has reportedly operated at design output with high mechanical reliability following its installation and startup in mid-summer 1972.

THE LAS VIRGENES PROJECT

A single-stage, covered-tank oxygenation system has been designed by the Cosmodyne Division of Cordon International Corporation. The system, given the name "SIMPLOX" and shown schematically in Figure 6, utilizes an inflated dome-type cover to contain the oxygen-rich atmosphere over the aerator. This concept is intended primarily for upgrading existing air activated sludge plants with a minimum capital expenditure by utilizing conventional air blowers and coarse-bubble air diffusers to recirculate oxygen gas. Air blowers used in this service must be corrosion proofed and otherwise modified to be compatible with oxygen gas. Virgin oxygen gas is introduced to the aerator through a fine-bubble sparger located on the tank bottom and on the opposite side wall from the conventional air diffusers. Power required for oxygen dissolution is greater for the "SIMPLOX" process than for the multi-stage systems because: (1) the equipment used for transferring oxygen is modified air aeration equipment and not specifically tailored to oxygen gas kinetics and (2) the gas phase above the mixed liquor is completely mixed and assumes the same oxygen composition as the exhaust gas stream; thus, the driving force for dissolving oxygen in wastewater is less than in the lead stages of multi-stage aerators. However, capital costs for converting an existing aerator from air to oxygen service should be significantly less with the "SIMPLOX" approach because staging baffles and multiple oxygen dissolution equipment assemblies are not required. Since the gas phase is completely mixed, exhaust oxygen, carbon dioxide, and inert gases can be bled from any point of the inflated dome and any of several activated sludge flow configurations, including plug flow, complete mix, and step aeration, can be used as desired.

A research and development grant was awarded to the Las Virgenes (California) Municipal Water District (a suburb of Los Angeles) in June 1971 to evaluate the "SIMPLOX" system at its Tapia Water Reclamation Facility. The experimental program concluded on September 10, 1973. The District contributed $62,000 in support of the project, supplementing the $186,000 EPA grant. An empty nominal one mgd train was available for the oxygen study because of a recent expansion at the Tapia facility. The manner in which the oxygen system was incorporated into this existing train is shown in plan view in Figure 7.

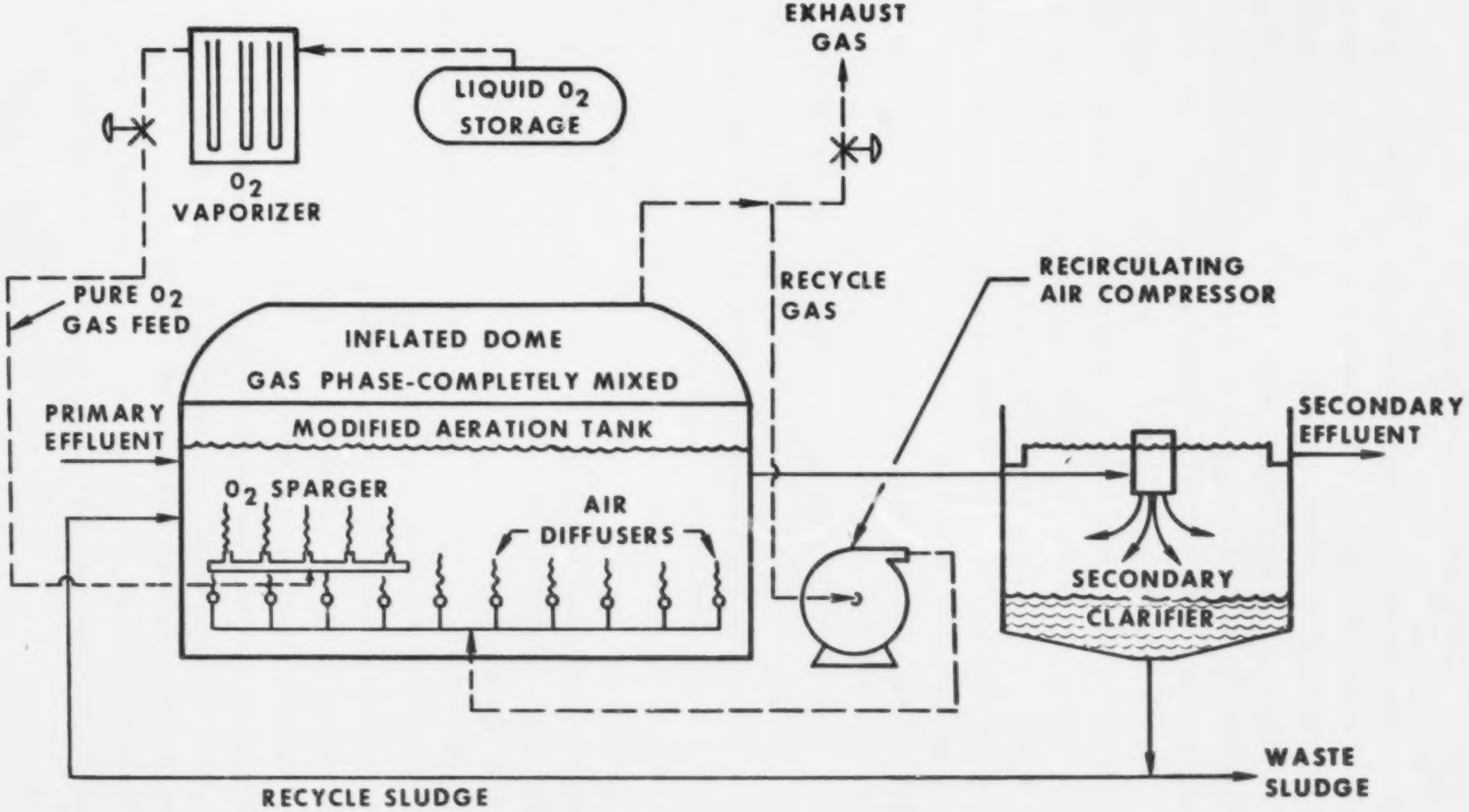

FIGURE 6. SCHEMATIC DIAGRAM OF DIFFUSED AIR AERATION SYSTEM MODIFIED TO RECIRCULATE OXYGEN GAS, LAS VIRGENES PROJECT

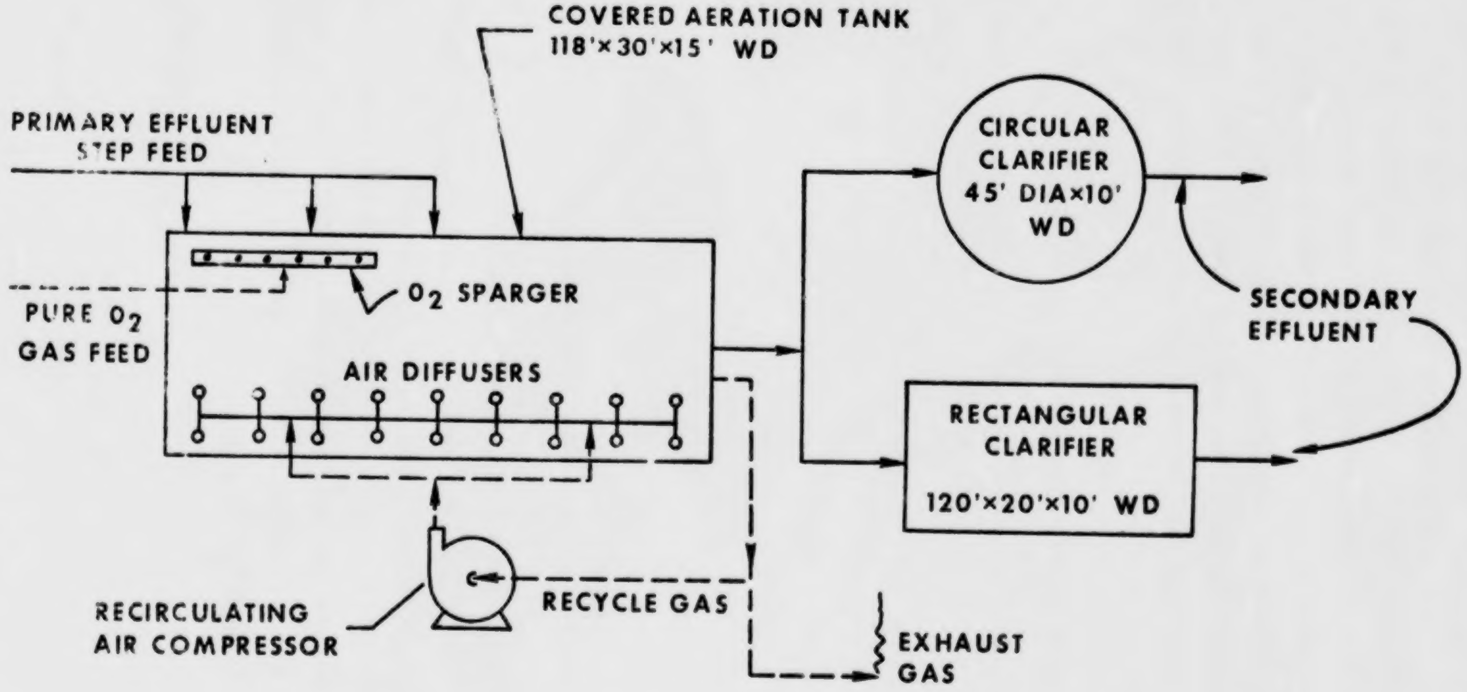

FIGURE 7. FLOW DIAGRAM FOR LAS VIRGENES OXYGENATION SYSTEM

The schedule followed during the experimental program for the project is outlined in Table 10. The range of aerator loadings examined was not as broad as at Newtown Creek due to influent flow limitations and a weaker aerator feed (primary effluent at Las Virgenes, raw sewage at Newtown Creek). The experimental program consisted of seven phases characterized by increasing flow and system loadings. To effect a more pronounced increase in aerator loading, only 45 percent of the available aerator volume was utilized in the last five phases. This was accomplished via the installation of a temporary bulkhead across the width of the aeration tank after Phase 2.

TABLE 10. EXPERIMENTAL SCHEDULE FOR LAS VIRGENES PROJECT (4/25/72-9/10/73)

Phase	Dates	Influent Flow (mgd)	% of Aerator In Use	No. of Clarifiers In Use
1	4/25/72 - 7/31/72	1.0	100	1
2	9/11/72 -11/13/72	2.0	100	2
3	1/22/73 - 3/8/73	1.0	45	1
4	3/9/73 - 4/3/73	1.13	45	2
5	4/4/73 - 4/30/73	1.3	45	2
6	5/1/73 - 5/14/73	1.54	45	2
7	5/15/73 - 9/10/73	1.85	45	2

System performance for the Las Virgenes project is summarized in Table 11. Tables 12, 13, and 14 summarize, respectively, system sludge characteristics, aerator loadings, and secondary clarifier loadings. Project data and information are presented in more thorough fashion in the final grant report. This report, currently being reviewed by EPA, is scheduled to be available for distribution by the end of the first quarter of 1975

TABLE 11. PERFORMANCE SUMMARY FOR LAS VIRGENES
(4/25/72 - 9/10/73)

	Phase						
	1	2	3	4	5	6	7
Total BOD_5 In (mg/l)*	82	69	79	107	115	103	95
Total BOD_5 Out (mg/l)	2	4	2	5	9	9	10
% Removed	97	94	97	95	92	91	89
Total COD In (mg/l)*	153	136	170	218	262	242	238
Total COD Out (mg/l)	35	35	29	35	37	40	50
% Removed	77	74	83	84	86	83	79
Soluble COD In (mg/l)*	58	43	76	93	101	101	100
Soluble COD Out (mg/l)	16	19	23	26	31	31	32
% Removed	72	56	70	72	69	69	68
Susp. Solids In (mg/l)*	73	67	39	53	63	59	44
Susp. Solids Out (mg/l)	9	7	4	7	5	4	6
% Removed	88	90	90	87	92	93	86
Turbidity Out (JTU)	2	3	2	3	2	2	3
NH_3-N In (mg/l)*	13.0	6.8	10.7	14.2	15.6	15.8	15.6
NH_3-N Out (mg/l)	0.4	0.1	0.2	4.1	4.8	2.8	3.1
% Removed	97	99	98	71	69	82	80
NO_3-N Out (mg/l)	16.2	15.3	8.8	6.9	5.6	7.5	8.0

*Concentrations shown are for primary effluent feed to oxygen aerator.

TABLE 12. AVERAGE SYSTEM SLUDGE CHARACTERISTICS FOR LAS VIRGENES (4/25/72-9/10/73)

Phase	MLSS (mg/l)	MLVSS (mg/l)	Return Sludge Flow (% of Q)	Return Sludge TSS (mg/l)	SVI (ml/gram)	SRT (days)
1	3,700	2,950	30	14,325	99	79
2	3,750	3,050	30	13,295	179	68
3	3,815	2,950	32	12,890	175	46
4	3,570	2,715	32	9,230	200	30
5	3,050	2,485	40	7,105	247	12
6	2,595	2,170	39	6,705	191	9
7	2,535	2,115	40	8,350	117	12

TABLE 13. AVERAGE AERATOR LOADINGS FOR LAS VIRGENES (4/25/72-9/10/73)

Phase	Detention Time -Based on Q- (hr)	F/M Loading $\left(\frac{\text{lb BOD}_5\text{/day}}{\text{lb MLVSS}}\right)$	Volumetric Organic Loading $\left(\frac{\text{lb BOD}_5\text{/day}}{1{,}000 \text{ ft}^3}\right)$
1	9.56	0.07	13
2	4.78	0.11	22
3	4.30	0.15	27
4	3.81	0.24	42
5	3.31	0.33	52
6	2.79	0.41	56
7	2.32	0.46	62

TABLE 14. AVERAGE SECONDARY CLARIFIER LOADINGS FOR LAS VIRGENES (4/25/72-9/10/73)

Phase	Surface Overflow Rate (gpd/ft^2)	Mass Loading $\left(\frac{lb\ TSS/ft^2}{day}\right)$
1	417	16.7
2	501	20.4
3	417	17.5
4	283	11.1
5	326	11.6
6	386	11.6
7	464	13.7

A cursory review of Table 11 reveals that effluent quality for the entire Las Virgenes project was superb and surpassed that observed at Newtown Creek. This can be attributed to three factors: (1) the lower aerator organic loadings which permitted a high degree of COD insolubilization, (2) the very conservative secondary clarifier surface and mass loadings which promoted highly effective solids capture, and (3) the lack of any significant industrial waste contributions. A major objective of wastewater treatment in the Las Virgenes District is the production of an ultra high quality secondary effluent after chlorination suitable for agricultural reuse. The thrust of this project, therefore, was geared not so much to maximizing system loadings (as was the case at Newtown Creek) as maintaining truly superb quality effluent and determining the effect of a relative conservative progression in system loadings on single-stage nitrification. As shown in Table 15, virtually complete nitrification was observed with F/M loadings between 0.07 and 0.15 lb BOD_5/day/lb MLVSS. For F/M loadings between 0.24 and 0.46 lb BOD_5/day/lb MLVSS, nitrification was only 69-82 percent complete. Lower wastewater temperatures may also have played a role in the decreased nitrification of the latter four phases.

TABLE 15. EFFECT OF ORGANIC LOADING AND WASTEWATER TEMPERATURE ON NITRIFICATION AT LAS VIRGENES (4/25/72-9/10/73)

Phase	F/M $\left(\frac{\text{lb BOD}_5\text{/day}}{\text{lb MLVSS}}\right)$	SRT (days)	Wastewater Temp. Range (°F)	% NH_3-N Removed	Fin. Eff. NO_3-N (mg/l)
1	0.07	79	70-77	97	16.2
2	0.11	68	73-79	99	15.3
3	0.15	46	65-67	98	8.8
4	0.24	30	65-67	71	6.9
5	0.33	12	67-70	69	5.6
6	0.41	9	68-71	82	7.5
7	0.46	12	70-75	80	8.0

Another major goal of the Las Virgenes staff was to minimize excess biological sludge production as much as possible. This goal probably led to the most significant problem area encountered on the project, a very evident bulking sludge. No sludge was intentionally wasted from the system during Phases 1 and 2. Wastage of suspended solids in the final effluent and final clarifier skimmings was sufficient to balance net system biomass growth at the low F/M loadings employed. Resulting SRT's were as high as 79 days and the SVI climbed to a level near 200 ml/gram. Despite the instigation of a scheduled wasting program in Phase 3, the sludge continued to bulk and the SVI climbed even higher. It was not until Phase 7 at an SRT of 12 days and an F/M loading of 0.46 lb BOD_5/day/lb MLVSS that a significant drop in SVI occurred. The bulking sludge condition is attributed here to a combination of Sphaerotilus filamentous development due to the inordinately high SRT's and the accumulation of other poor settling debris in the floc matrices. It was only because of the low clarifier loadings that efficient overall performance was sustained. Sludge blanket levels frequently rose to within a few feet of the clarifier weirs. The Las Virgenes experience illustrates the potential operating difficulties that can and probably will occur at very low oxygen system loading rates.

One definite conclusion reached during the project is that the inflated tent (dome) concept is not suitable for permanent installation. New leaks developed repeatedly in the polyvinyl material due to separation of the tent/tank interface, abrasion against the tent support structure during high winds, and bullets from pranksters' guns. The gas leak problem made accurate oxygen consumption monitoring impossible, and during the latter higher loading phases, the leaks became sufficiently frequent and large that it was extremely difficult to maintain a mixed liquor D.O. above 1-2 mg/l. The rationale for using an inflated dome in lieu of a flat cover on this research project was to permit access to the tank interior, a procedure effectively utilized on several occasions. A permanent installation would probably require a flat, more rigid cover for longevity and minimization of leaks.

The Cosmodyne Division of Cordon International has not attempted to establish a proprietary position with respect to the "SIMPLOX" system. Notwithstanding the attractive capital cost features of this oxygen dissolution concept for upgrading existing air-activated sludge plants, without the support of a proprietary interest and an aggressive marketing effort, utilization of this process in treatment plant construction will most likely proceed at a much slower rate than with other oxygen processes.

THE FMC PROJECT

The FMC Corporation has developed a unique fine-bubble diffuser capable of producing uniform oxygen bubbles of less than 0.2 mm in diameter. The diffuser works on the shear principle of passing a high velocity liquid stream at right angles to oxygen bubbles discharging into a vertical slot from capillary tubes. Oxygen gas is introduced to the capillary tubes at 30 psi pressure. A graph provided by FMC showing water depth required for complete dissolution of varying size oxygen gas bubbles is reprinted in Figure 8.* The large effect of a relatively small change in bubble size on the water depth required for 100 percent dissolution is readily evident. For a bubble diameter of 0.20 mm, a 17.5 foot deep tank would be required. The required depth decreases to 8.5 feet for a 0.15 mm diameter bubble.

One of the many potential applications for this diffuser is in an open-

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*This graph was prepared using tap water. Dissolution characteristics for various size oxygen gas bubbles may and probably do differ for a wastewater undergoing biological treatment.

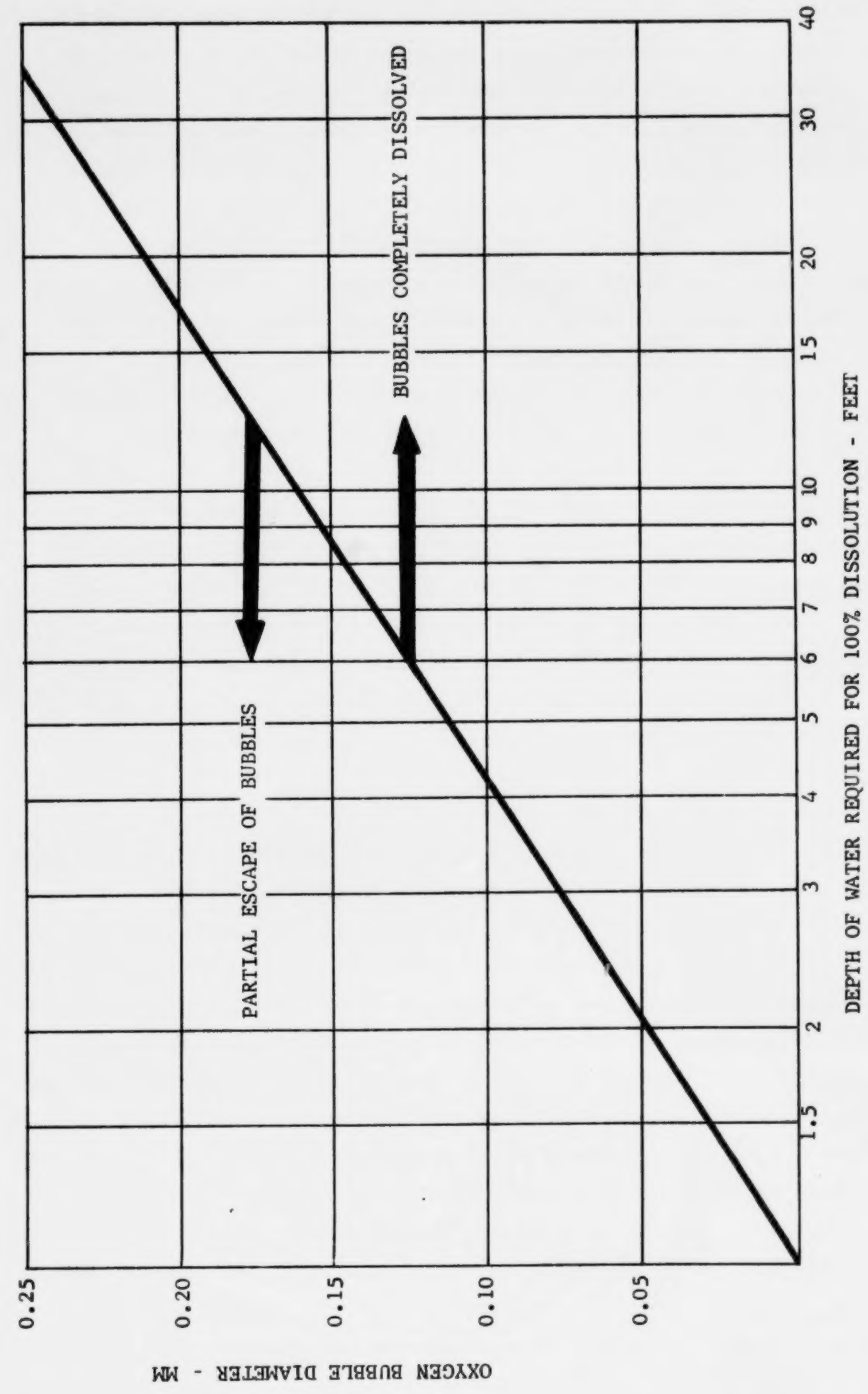

FIGURE 8. OXYGEN GAS BUBBLE DIAMETER VS. WATER DEPTH FOR COMPLETE DISSOLUTION

tank oxygen-activated sludge process. To evaluate the feasibility of an open-tank oxygenation approach, a research and development grant for $142,000 was awarded to FMC in September 1972 for a nominal 30 gpm pilot plant study. The firm is contributing over $75,000 of their funds to the project. The pilot plant has been installed on the grounds of the Englewood, Colorado (suburb of Denver), trickling filter plant and receives a feed stream of primary effluent from that plant. Pilot plant configuration and dimensions are shown in Figure 9. The aeration tank is provided with two baffles to approximate a plug flow (three-stage) condition. Diffusers are located in each of the stages. Mixed liquor is recirculated through the diffusers by low head centrifugal pumps. Pump suction is taken near the liquid surface to promote mixing and tank turnover. Throttling of the oxygen feed is accomplished automatically by D.O. sensing and control.

Major points of research interest in the project are: (1) oxygen utilization efficiency in an open-tank setting, (2) oxygen feed control response based on a D.O. monitoring approach, (3) mixed liquor recirculation rates and power requirements, (4) diffuser self cleansing (non-clogging) capabilities, and (5) shearing effect, if any, on mixed liquor particles caused by continuous recirculation through the pumps and diffusers. In the event that floc disruption did occur, a short detention biological reflocculation tank (gentle mixing, no chemicals) was interposed between the aerator and secondary clarifier. Two aspects of system design which cannot be adequately defined at the scale of this pilot plant study are diffuser mixing characteristics and additional mixing requirements, if any, for large aeration tanks. This task is being addressed by FMC in deep tank tests using tap water at both the firm's Englewood and Santa Clara laboratories.

Pilot plant fabrication was completed in late June 1973. System startup required the first 20 days of July. The experimental program which followed was divided into eight phases and is outlined in Table 16. Performance data for the four highest flow phases (Phase 4 through 7) are summarized in Table 17. Sludge characteristics, aerator loadings, and secondary clarifier loadings for the same four phases are presented in Tables 18, 19 and 20 respectively. Data for the first three conservative load phases as well as Phase 8 which was still in progress at the date of this writing will be included by FMC in the final project report. Availability of this report is expected by mid-1975.

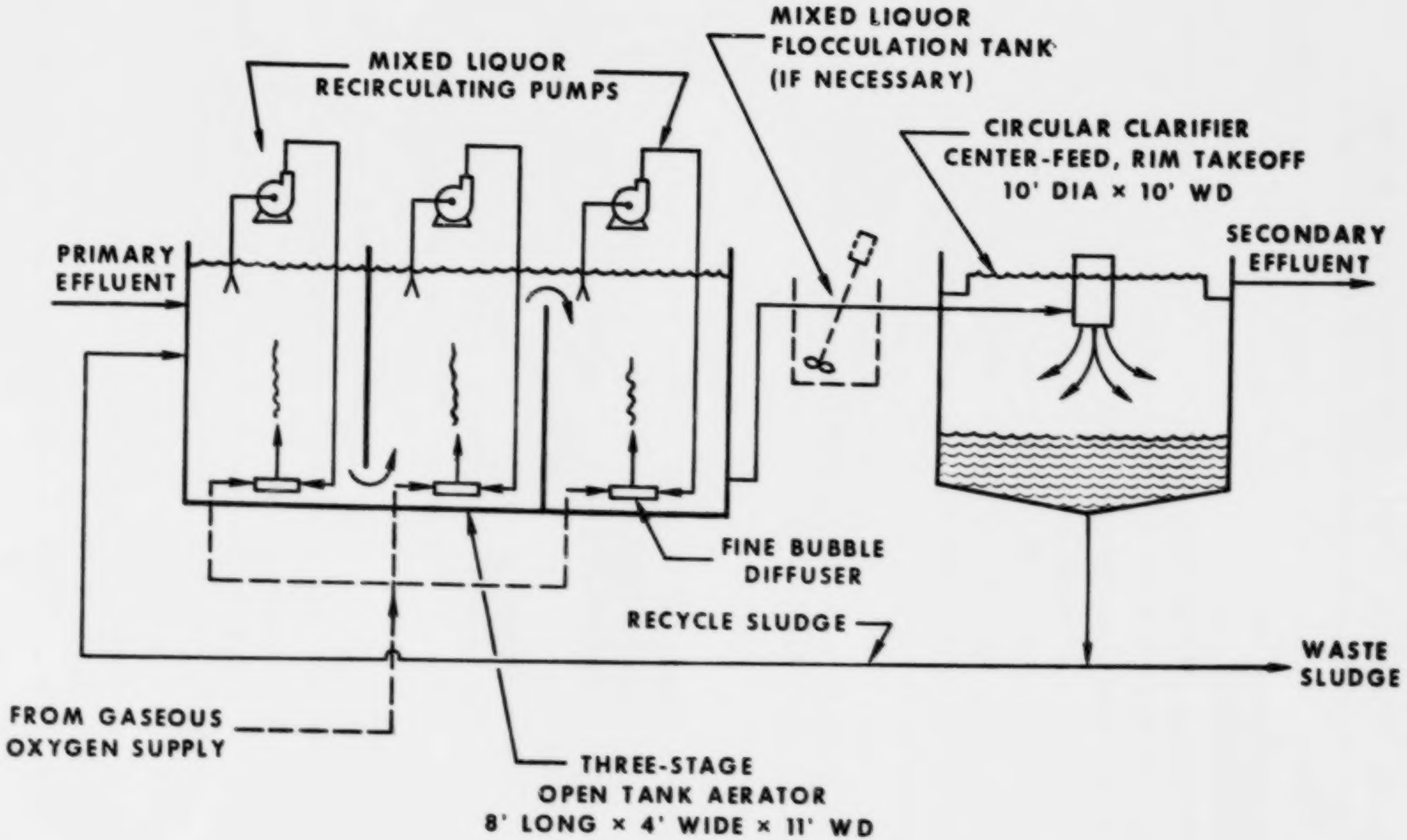

FIGURE 9. FMC OPEN-TANK OXYGENATION PILOT SYSTEM

TABLE 16. PLANNED EXPERIMENTAL PROGRAM FOR FMC PROJECT

Phase	Dates	Influent Flow Condition	No. of Clarifiers in use
1	7/21/73 - 9/5/73	10 gpm (Constant)	1
2	9/6/73 - 9/30/73	10 gpm (Diurnal)	1
3	12/6/73 - 1/28/74	15 gpm (Constant)	1
4	4/8/74 - 4/30/74	25 gpm (Constant)	2
5	5/1/74 - 5/31/74	35 gpm (Constant	2
6	6/1/74 - 6/30/74	30 gpm (Diurnal)	2
7	7/1/74 - 7/31/74	20 gpm (Diurnal)	1
8	10/1/74 -10/31/74	15 gpm (Constant)	1

TABLE 17. PERFORMANCE SUMMARY FOR FMC PROJECT (4/8/74 - 7/31/74)

	Phase			
	4	5	6	7
Total BOD_5 In (mg/l)*	153	159	180	208
Total BOD_5 Out (mg/l)	13	18	16	16
% Removed	92	89	91	92
Total COD In (mg/l)*	332	315	259	322
Total COD Out (mg/l)	95	74	57	61
% Removed	71	77	78	81
Suspended Solids In (mg/l)*	110	85	85	115
Suspended Solids Out (mg/l)	13	15	12	15
% Removed	88	82	86	87
Turbidity Out (JTU)	6	6	4	5
Sewage Temperature (°F)	57	62	67	71

* Concentrations shown are for primary effluent feed to oxygen aerator.

TABLE 18. AVERAGE SYSTEM SLUDGE CHARACTERISTICS FOR FMC PROJECT (4/8/74 - 7/31/74)

Phase	MLSS (mg/l)	MLVSS (mg/l)	Return Sludge Flow (% of Q)	Return Sludge TSS (mg/l)	SVI (ml/gram)	SRT (days)
4	5,120	4,010	11,585	60	71	2.0
5	4,030	3,435	10,485	52	70	1.5
6	4,745	3,860	12,220	57	67	2.1
7	3,960	3,365	10,850	50	73	2.6

TABLE 19. AVERAGE AERATOR LOADINGS FOR FMC PROJECT (4/8/74-7/31/74)

Phase	Detention Time -Based on Q- (hr)	F/M Loading $\left(\frac{\text{lb BOD}_5\text{/day}}{\text{lb MLVSS}}\right)$	Volumetric Organic Loading $\left(\frac{\text{lb BOD}_5\text{/day}}{1{,}000\ \text{ft}^3}\right)$
4	1.32	0.69	173
5	0.94	1.17	253
6	1.10	1.01	244
7	1.65	0.91	190

TABLE 20. AVERAGE SECONDARY CLARIFIER LOADINGS FOR FMC PROJECT (4/8/74-7/31/74)

Phase	Surface Overflow Rate (gpd/ft^2)*	Mass Loading $\left(\frac{\text{lb TSS/ft}^2}{\text{day}}\right)$*
4	514	35
5	720	37
6	617	38
7	823	41

* Excludes influent center-well annular area which = 9.1% of total clarifier surface area.

As shown in Table 17, BOD_5 and suspended solids removals during the high loading conditions of Phases 4, 5, 6, and 7 were excellent. One of the significant observations forthcoming from the project was that the feared disruption of sludge settling properties due to floc shearing as the mixed liquor was continually recirculated through the centrifugal pumps and diffusers did not materialize. SVI for the above four phases averaged a very acceptable 70 ml/gram. A highly concentrated float (4-6 percent TSS) approximately six inches thick quickly developed on the surface of the aerator after startup. This combination aeration/flotation effect was anticipated in light of the fine bubbles created by the oxygen diffusers. It was found that the thickness of the float can be controlled by adjusting the elevation at which the mixed liquor recirculation suction is taken. FMC personnel believe this feature offers a potential economically attractive alternative location for extracting waste sludge from an activated sludge system.

In a full-scale embodiment of this open-tank oxygen concept, mixed liquor recirculation would not be accomplished by centrifugal pumps. Instead, FMC envisions a propeller-type pump mounted inside a downcomer draft tube. The draft tube in turn is to be connected to a pipe header containing many gas bar diffusers. The whole assembly will rest on the aeration tank floor and will be prevented from moving sideways by lateral catwalks which are tied into the top of the draft tubes. Elevation, plan, and side views of a full-scale aerator assembly as currently proposed are pictured in Figure 10. A perspective view of a typical aeration tank containing several of these assemblies and the resulting fluid mixing pattern are shown in Figure 11. This system appears to possess the essential ingredients for significantly impacting the wastewater treatment construction industry.

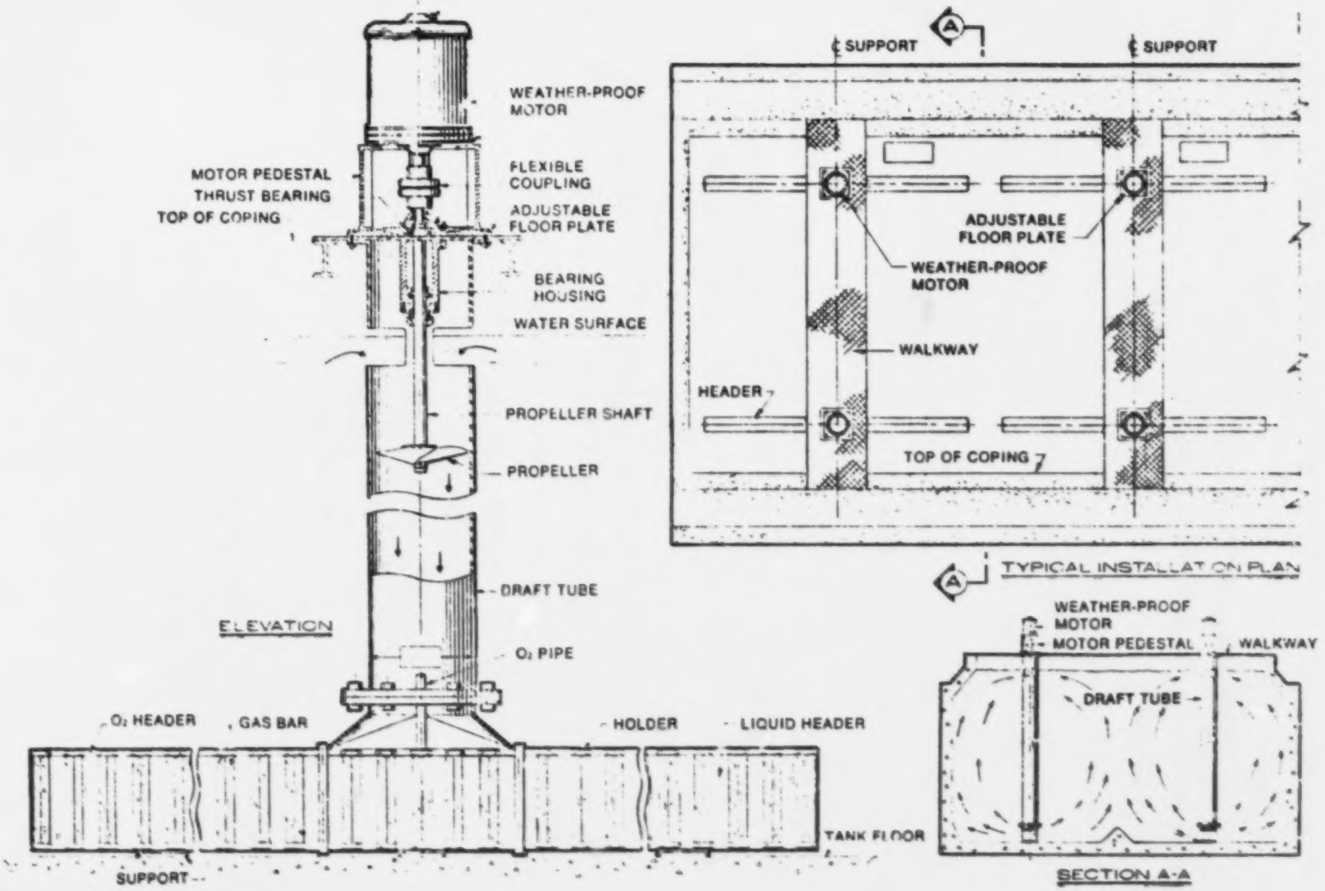

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FIGURE 10. ELEVATION, PLAN, AND SIDE VIEWS OF ENVISIONED FULL-SCALE EMBODIMENT OF FMC "MAROX" OPEN-TANK OXYGENATION SYSTEM

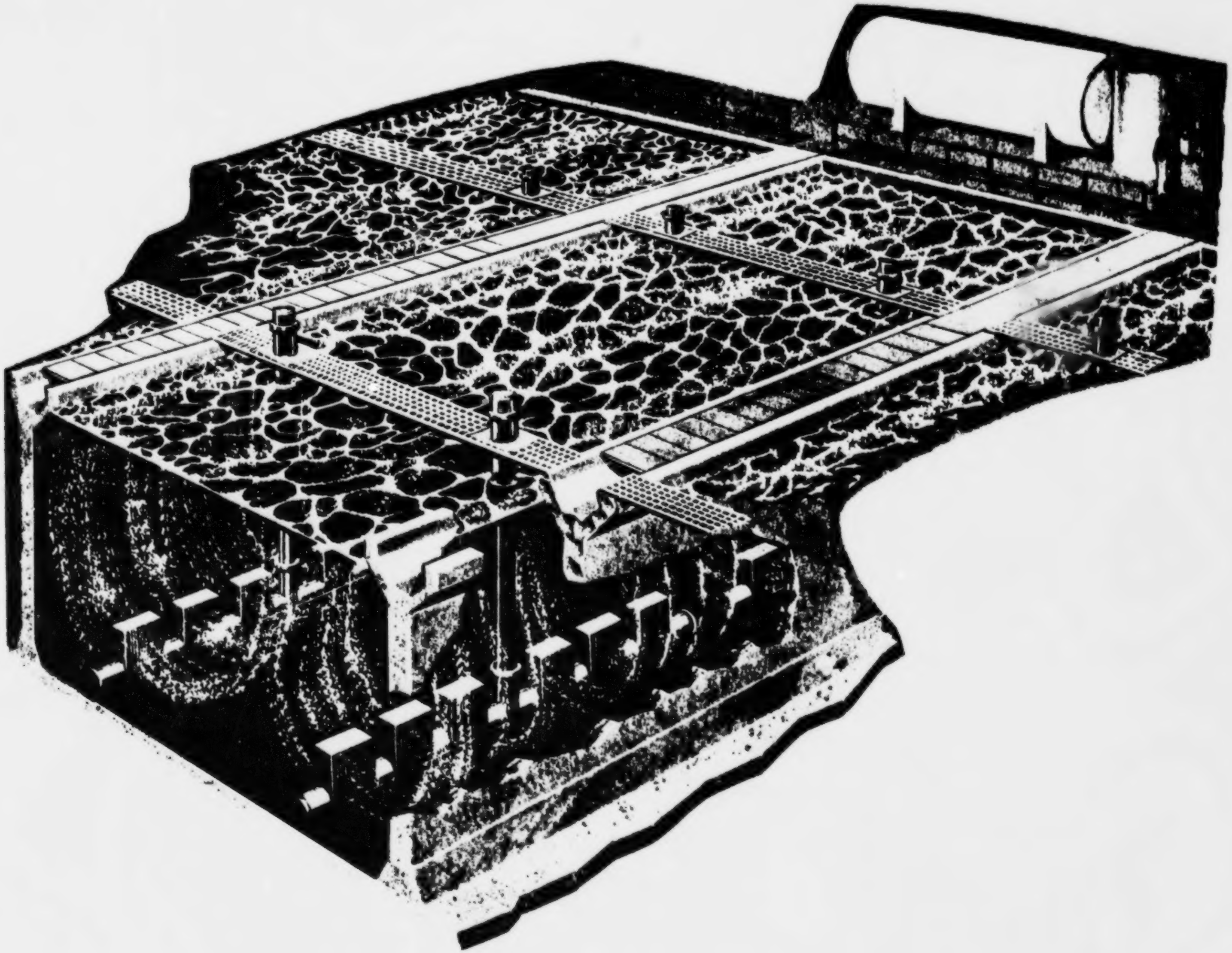


FIGURE 11. PERSPECTIVE VIEW OF ENVISIONED
FULL-SCALE "MAROX" OPEN-TANK OXYGEN SYSTEM

THE BLUE PLAINS PROJECT

A multi-stage, covered tank oxygenation pilot system of Union Carbide design (see Figure 12) was operated continuously from June 1970 through September 1972 at the Joint EPA/District of Columbia (Blue Plains) Pilot Plant. Nominal design throughput for the system was 70 gpm (100,000 gpd). The results generated in over two years of work (believed to be the single longest continuous oxygenation pilot plant study on record) were extensively reported at the 1972 Water Pollution Control Federation Conference (Stamberg, et al., 1972). For a detailed summary of monthly operating data, the reader is referred to the upcoming publication of this paper in the Federation Journal. Discussion of the project here is limited to generalized results and observations.

The oxygen system was operated over a wide range of SRT's from 1.3 to 13.0 days. However, on District of Columbia (D.C.) primary effluent, filamentous organisms propagate rapidly with either oxygen or air if the SRT is held below approximately five days for any extended period of time, producing a bulking sludge with greatly retarded settling rates. Consequently, the majority of the Blue Plains operation has been intentionally restricted to SRT's greater than five days. A technique devised by project staff personnel of reducing the incoming flow and twin dosing the sludge recycle stream with 200 mg/1 of hydrogen peroxide (based on influent flow) for 24-hour periods at a one-week interval proved to be an effective method for purging entrenched filamentous bacterial growths from an activated sludge system. The technique provides lasting benefit only if subsequent F/M loadings are adjusted to maintain an SRT outside the critical filamentous growth range. The conditions under which filamentous cultures propagate and flourish are unique to each wastewater and location. Some plants can operate in any desired loading range without encountering filamentous problems. Oxygen mixed liquor at Blue Plains was normally well bioflocculated and essentially free of fragmented debris between discrete particles.

Above an SRT of five days, average system F/M loadings remained in the range of 0.27-0.50 lb BOD_5/day/1b MLVSS. On those few occasions when

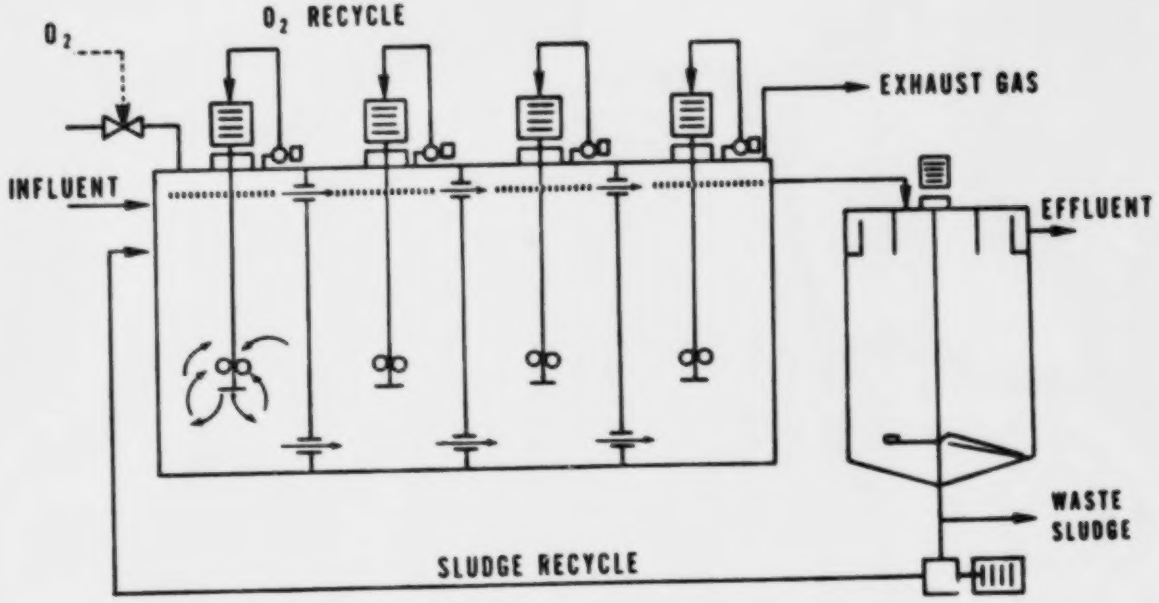


FIGURE 12. SCHEMATIC DIAGRAM OF BLUE PLAINS OXYGENATION SYSTEM[a]

[a] Reprinted with permission (Stamberg, et al., 1972)

the system was operated at an SRT less than five days, F/M loadings rose to levels as high as 1.0 lb BOD_5/day/lb MLVSS. Corresponding average volumetric organic loadings at an SRT above five days ranged from 57-185 lb BOD_5/day/1000 ft^3. Aerator detention times (based on Q) were varied between 1.5 and 2.8 hours throughout the two-year+ period. For all loadings investigated, BOD_5 insolubilization was virtually complete. Effluent soluble BOD_5 residuals were never greater than 5 mg/l and consistently averaged 2-3 mg/l. Total BOD_5 and suspended solids removal were a direct function of clarifier performance. Effluent COD and TOC concentrations typically ranged from 35-60 and 15-20 mg/l, respectively.

During the spring periods of rising wastewater temperature, nitrification was established more slowly in the Blue Plains single sludge oxygen system than in a parallel conventional single sludge air aerated pilot system probably due to the lower mixed liquor pH inherent in operation of a covered biological reactor. Once established, however, substantial nitrification was exhibited by the oxygen system during warm weather. With decreasing wastewater temperature in the fall, deterioration of nitrication was directly related to SRT. At an SRT of 9.0 days and a wastewater temperature of 63°F, at least partial nitrification was sustained. However, once the wastewater temperature decreased to about 60°F, no nitrification was observed in the Blue Plains oxygen system up to an SRT of 13.0 days.

Phosphorus removal experiments were conducted by adding aluminum sulfate (alum) directly to the oxygen mixed liquor. At an Al^{+++}/P weight ratio of 1.4/1.0, phosphorus removal averaged 80% with total and soluble phosphorus residuals of 1.8 and 1.6 mg/l (as P), respectively. Increasing the alum dose to an Al^{+++}/P weight ratio of 1.8/1.0 decreased total and soluble residuals to 0.62 and 0.53 mg/l (as P), respectively, but it also lowered mixed liquor pH from 6.5 to 6.0. At the lowered pH, the oxygen biota eventually dispersed and the experiments were discontinued. For low alkalinity wastewaters such as the District of Columbia's, pH control may be necessary to achieve efficient (90% or greater) phosphorus removal when acidic metallic salts are added directly to oxygen-activated sludge mixed liquor.

Oxygen clarifier performance at Blue Plains and its effect on total system operation are addressed in a later section. Continued experiments only recently completed at Blue Plains included evaluation of oxygen in a step aeration flow regime and examination of the nitrification kinetics of a second-stage oxygen system operating on full-scale D.C. modified aeration effluent feed. Reports on these activities are in preparation.

THE BUREAU OF RECLAMATION PROJECT

The Bureau of Reclamation's Engineering and Research Center in Denver, under an interagency agreement with EPA, has recently completed the second year of a three-year project to test many different materials of construction to evaluate their suitability for use with oxygen aeration wastewater treatment systems. The materials being tested include three different types of concrete, twelve different metals, and eleven protective coatings, linings, joint sealers, and gaskets.

The materials are being exposed for varying lengths of time to oxygen-rich mixed liquor, oxygen-rich vapor above the mixed liquor, and to the interface between the two phases and then withdrawn for examination. Oxygen reactors being utilized for these tests include Las Virgenes; Speedway, Indiana; and Fairfax County, Virginia. Interim results are available by writing to EPA, Office of Environmental Engineering, Washington, D.C. (20460).

CRITICAL PROCESS PARAMETERS

Certain process parameters are vital to the successful operation and economic attractiveness of all waste treatment processes. For oxygen aeration systems, four of these process parameters are oxygen utilization and consumption, sludge production, power consumption, and biological performance versus biomass loading. Available data for the projects described above are summarized below for each of the four parameters.

Oxygen Utilization and Consumption

A misconception which seems to have accompanied the development of the oxygenation processes is that oxygen gas possesses mystical qualities and can oxidize organics and ammonia nitrogen with less oxygen consumption than air systems. In reality, of course, the same amount of oxygen is required to oxidize a given amount of organic carbon to carbon dioxide and water or a given amount of ammonia nitrogen to nitrate nitrogen regardless of the source of oxygen or the method in which it is delivered to a biological system.

The conventional method of calculating oxygen consumption in a covered-tank oxygen system is to monitor inlet and exhaust gas flows and effluent D.O. and assume that all oxygen not accounted for was consumed. This method will not detect any gas leaks which may develop in and at the joints of the reactor cover. A second method which is sensitive to detecting sizeable leaks and can be used to check the gross accuracy of the oxygen metering equipment is an oxygen balance technique recommended by the Blue Plains staff (Stamberg, et al., 1972) and shown in Table 21. The method assumes that one pound of oxygen is consumed for every pound of COD destroyed (not to be confused with COD removed from the substrate) and that 4.57 pounds of oxygen are consumed for every pound of ammonia nitrogen converted to nitrate nitrogen. The method is reasonably accurate provided the wastewater does not contain certain industrial components which do not consume oxygen in a COD determination but will utilize oxygen in a biological system.

Oxygen utilization and supply data for Newtown Creek, Batavia, Blue Plains, and the FMC project are summarized in Table 22. Accurate measurement of oxygen utilization at Las Virgenes was hampered due to excessive gas leaks in the tent cover previously described. The table indicates a lack of informity for all three methods selected for indicating specific oxygen supply requirements. Generally, in the absence of nitrification, slightly more than one pound of oxygen should theoretically be supplied for each pound of COD destroyed. The Newtown Creek value of 0.8 for the period of September 17, 1972 through October 14, 1972 cannot possibly be correct and indicates probably either low inlet gas measurements or low waste sludge COD determinations. Of the three methods shown, designing oxygen supply requirements on the basis of anticipated BOD_5 removal is the least reliable. Since COD destroyed usually

TABLE 21. OXYGEN BALANCE METHODS

Method 1:		lb/mil gal Oxygen Supplied
	(-)	lb/mil gal Exhaust Oxygen
	(=)	lb/mil gal Oxygen Utilized
	(-)	lb/mil gal Secondary Effluent D.O.
	(=)	lb/mil gal Oxygen Consumed
Method 2:		lb/mil gal Aerator Influent COD
	(-)	lb/mil gal Secondary Effluent COD
	(-)	lb/mil gal Waste Sludge COD
	(=)	lb/mil gal COD Destroyed[a]
	(+)	lb/mil gal Nitrate Nitrogen Oxygen Demand[b]
	(+)	lb/mil gal Exhaust Oxygen
	(+)	lb/mil gal Secondary Effluent D.O.
	(=)	lb/mil gal Oxygen Supplied (Theoretically)

[a] Assumes 1 lb COD destroyed consumes 1 lb O_2.

[b] Assumes 1 lb NH_3-N converted to 1 lb NO_3-N consumes 4.57 lb O_2.

cannot be accurately predicted in advance, using an oxygen required/anticipated COD removal weight ratio of 0.60-0.75 and adding this to anticipated nitrification oxygen demand, if any, is probably the best technique available for sizing oxygen supply equipment. The effect of nitrification on oxygen supply and consumption is readily apparent at Blue Plains in May 1972. Generally, but not always, less oxygen was consumed per pound of BOD_5 removed as the F/M loading increased. A similar pattern was not evident for oxygen consumed per pound of COD removed.

TABLE 22. SUMMARY OF OXYGEN UTILIZATION AND SUPPLY

Plant	Metered O_2 Utilization	F/M $\left(\frac{\text{lb BOD}_5\text{/day}}{\text{lb MLVSS}}\right)$	$\frac{\text{lb } O_2 \text{ Supplied}}{\text{lb BOD}_5 \text{ Removed}}$	$\frac{\text{lb } O_2 \text{ Supplied}}{\text{lb COD Removed}}$	$\frac{\text{lb } O_2 \text{ Supplied}}{\text{lb COD Destroyed}}$
Newtown Creek					
Phase 1	>90	0.65	1.09	0.55	0.80[a]
Phase 2	>90	0.57	1.20	0.59	-
Phase 3	>90	0.57	1.39	0.65	-
Phase 4	>90	0.92	1.02	0.55	-
Phase 5	>90	1.19	0.88	0.72	-
Phase 6	>90	1.62	0.83	0.73	-
Phase 7	>90	2.44	0.79	0.61	-
Batavia					
5/12/69-11/10/69	93	0.59	0.94	0.60	-
9/1/70-11/30/70	92	0.87	1.36[b]	1.13[b]	-
Blue Plains					
May 1971[c]	97	0.97	1.04	0.60	1.09
May 1972[d]	97	0.36	2.09	1.03	1.23
FMC Project					
Phase 4	-	0.69	1.22	0.72	-
Phase 5	-	1.17	1.11	0.65	-
Phase 6	-	1.01	0.92	0.75	-
Phase 7	-	0.91	1.08	0.79	-

[a] Covers the segment of Phase 1 from 9/17/72-10/14/72 only.

[b] Values are high due to high clarifier loadings resulting in significant solids carryover and lowered BOD_5 and COD removals.

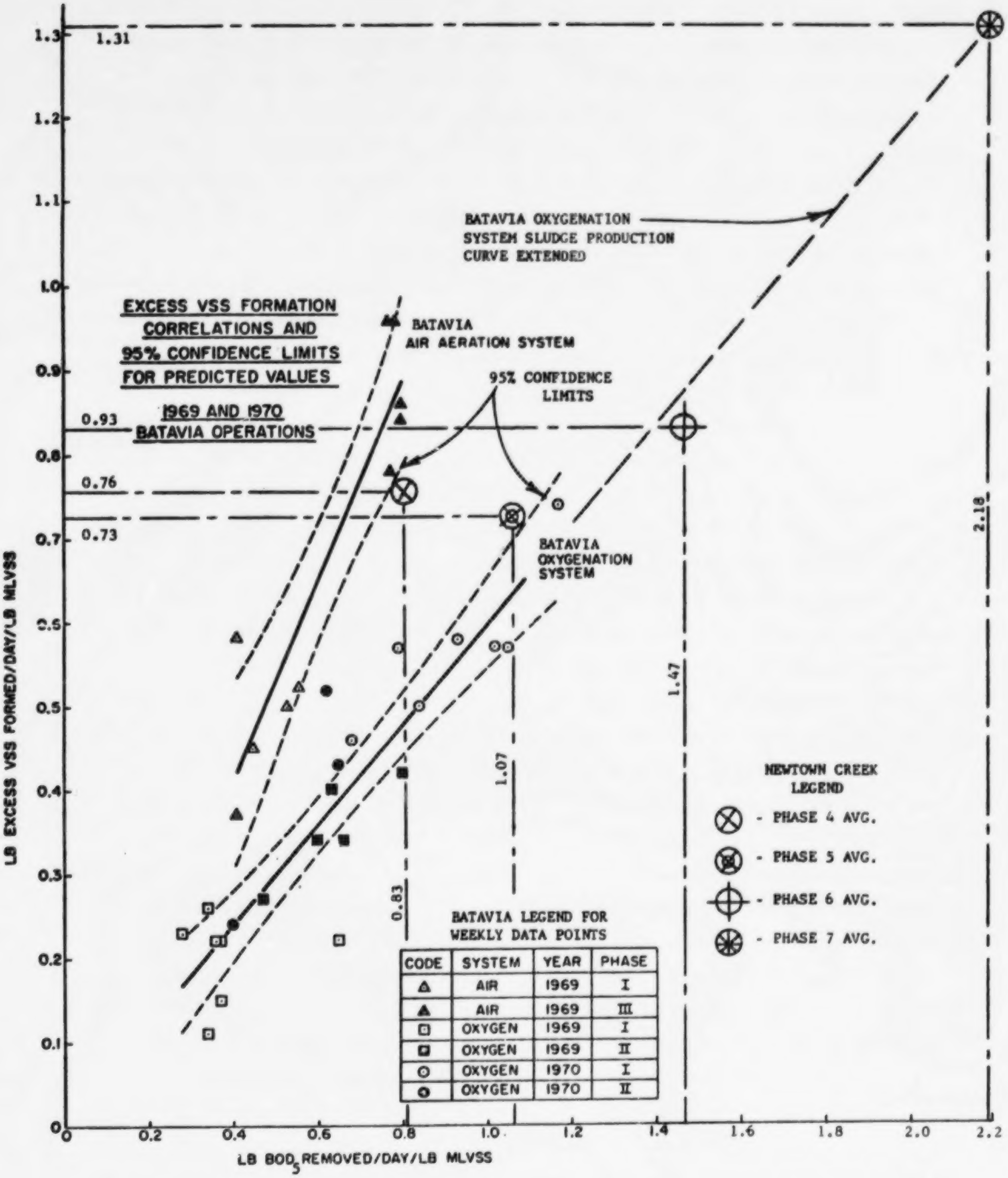
[c] No nitrification.

[d] Substantial nitrification.

Sludge Production

Available data indicate that oxygen systems may produce less excess biological sludge than air systems at comparable F/M loadings. The first indication was provided by the two Batavia projects as shown in Figure 13. In this figure, BOD_5 removed per day per unit of MLVSS is plotted in the conventional method against the inverse of SRT for both the air and oxygen trains. The curves reveal an approximate 50% reduction in favor of oxygen. Although both of these trains were operated in plug flow configuration, the comparison is most likely heavily biased toward oxygen because of the severe D.O. limitations under which the Batavia air reactor operated during these projects. If oxygen does produce less sludge, it is probably due to the high mixed liquor D.O. concentration maintained and the additional driving force it provides for increasing oxygen penetration into and stimulating aerobic activity within floc particle interiors. Air system sludge production data in which the mixed liquor is not devoid of D.O. for a lengthy section at the head of the aerator (in contrast to Batavia) will provide more meaningful future comparisons with oxygen sludge production data.

Reliable sludge production data are available from Newtown Creek for Phases 4-7. The data have been averaged for each phase and superimposed on the Batavia sludge production curve in Figure 13. Three of the four points plot reasonably close to the line of best fit (or that line extended) for the Batavia excess sludge production data, lending credibility to this projection of oxygen sludge production for raw wastewater feed. The fourth point (Phase 4), which represents the least loaded condition of the four phases plotted, falls considerably above the Batavia oxygen sludge production curve. Additional sludge production data generated by the Newtown Creek oxygen system during the extended operating period of September 1973 through April 1974 are presented in the previously mentioned New York City authored paper "Upgrading New York City Modified Aeration with Pure Oxygen." These additional data along with the data summarized in this report provide an accurate representation of oxygen system sludge production at Newtown Creek over a substantially broad range of loadings.



BATAVIA LEGEND FOR WEEKLY DATA POINTS

CODE	SYSTEM	YEAR	PHASE
△	AIR	1969	I
▲	AIR	1969	III
⊡	OXYGEN	1969	I
■	OXYGEN	1969	II
⊙	OXYGEN	1970	I
●	OXYGEN	1970	II

FIGURE 13. NEWTOWN CREEK EXCESS OXYGEN SLUDGE PRODUCTION DATA SUPERIMPOSED ON BATAVIA EXCESS SLUDGE PRODUCTION CORRELATIONS PLOT

Sludge production data for the Blue Plains oxygen system are compared with data from a parallel step aeration air pilot system operated on the same primary effluent feed in Figures 14 and 15 (Stamberg, et al., 1972). Figure 14 plots SRT versus F/M loading and indicates that less volatile mass under aeration was required with oxygen to reach any given SRT above six days. Blue Plains personnel attribute the increased activity of the oxygen volatile mass to maintaining mixed liquor D.O. between 4 and 8 mg/l and the minimization of sludge pockets and dead spots afforded by independently controlled mixing. Figure 14 can be manipulated to produce Figure 15 by multiplying F/M values by the corresponding SRT values and inverting the product to yield volatile solids produced per unit of BOD_5 applied and then replotting these new values against SRT. Figure 15 shows substantial reduction in sludge production with oxygen again above an SRT of six days. Below an SRT of about eight days, the step aeration system experienced soluble BOD_5 breakthrough and overall effluent quality was poorer than that of the oxygen system. Because of the different flow configurations utilized, sludge production information generated by these two systems cannot be used directly to derive conclusions regarding the relative sludge production rates of oxygen and air at comparable SRT's. The kinetics of step aeration dictate that it will experience maximum sludge production at a much higher SRT than a plug flow regime.

In an attempt to compare sludge production data from air and oxygen systems with similar operating conditions and with the further restriction that the air system is not oxygen transfer limited, data from the Hyperion Plant in Los Angeles (Smith, 1969) are plotted against the Blue Plains data in Figure 16 on a BOD_5 removed basis and Figure 17 on a COD removed basis. Hyperion is an air activated sludge plant with a consistent record of excellent performance and adequate mixed liquor D.O. Both Hyperion and Blue Plains data were collected on systems operated in a conventional plug flow mode on primary effluent feed. In Figure 16 a majority of the oxygen points fall below the Hyperion regression curve and Smith's computer program curve for Hyperion. In Figure 17 all but two of the oxygen points fall below the air curve, many of them well below. These two figures lend additional support to the position

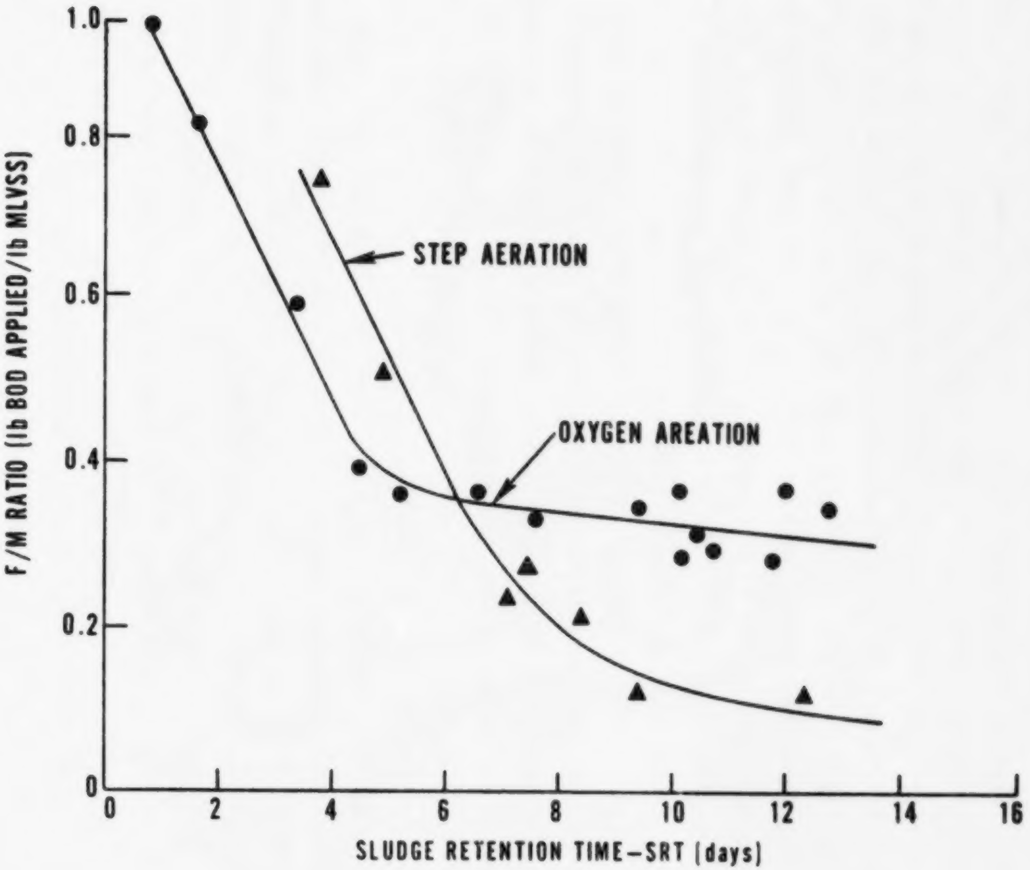


FIGURE 14. BIOLOGICAL ACTIVITY RELATIONSHIPS - BLUE PLAINS[a]

[a]Reprinted with permission (Stamberg, et al., 1972)

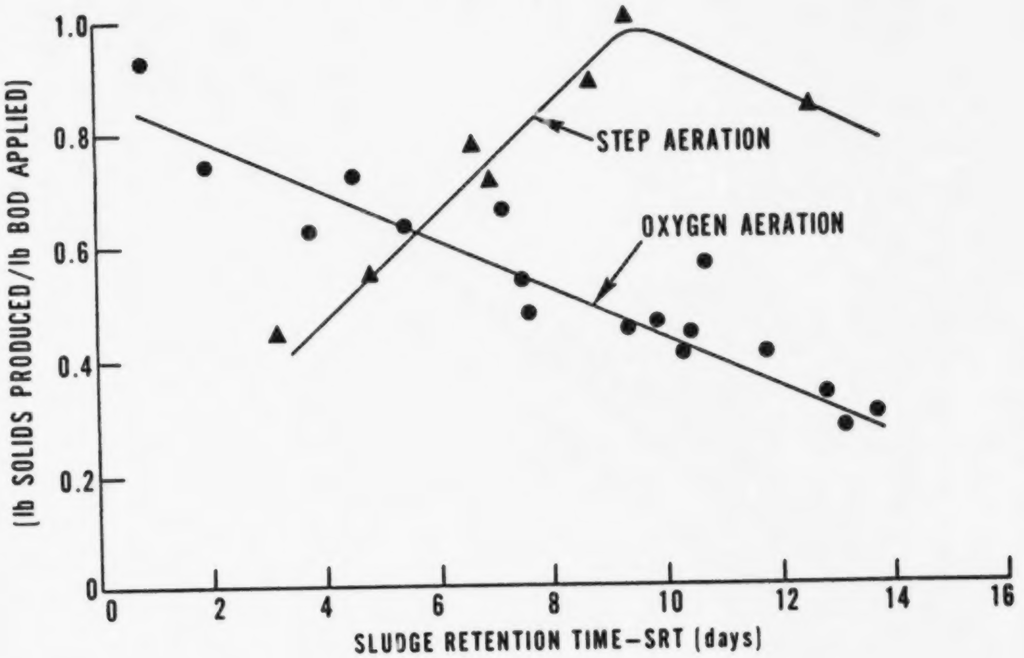


FIGURE 15. EXCESS BIOLOGICAL SLUDGE PRODUCTION - BLUE PLAINS[a]

[a] Reprinted with permission (Stamberg, et al., 1972)

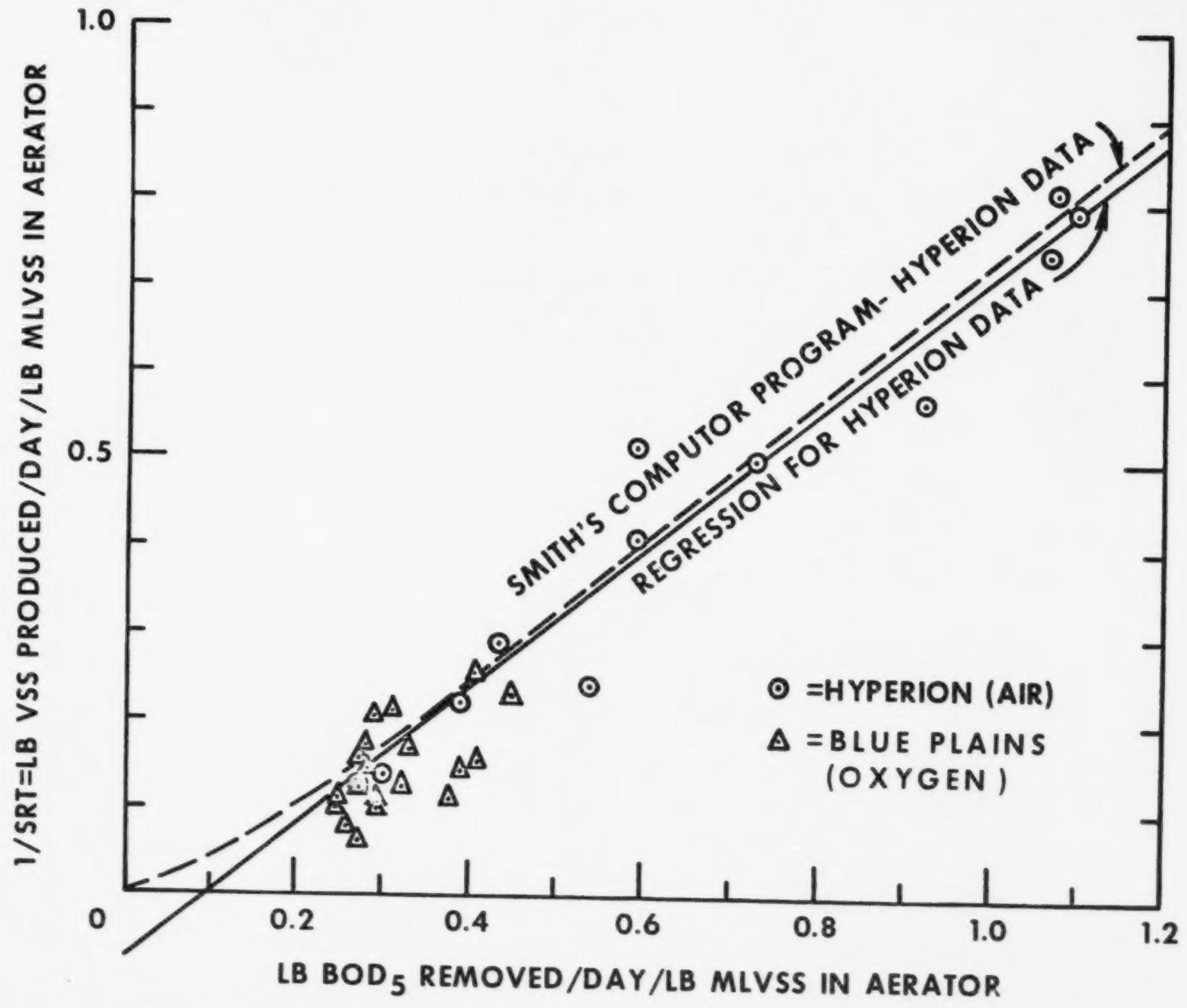


FIGURE 16. COMPARISON OF SLUDGE PRODUCTION FOR AIR AND OXYGEN SYSTEMS ON PRIMARY EFFLUENT FEED (BOD$_5$ REMOVED BASIS)

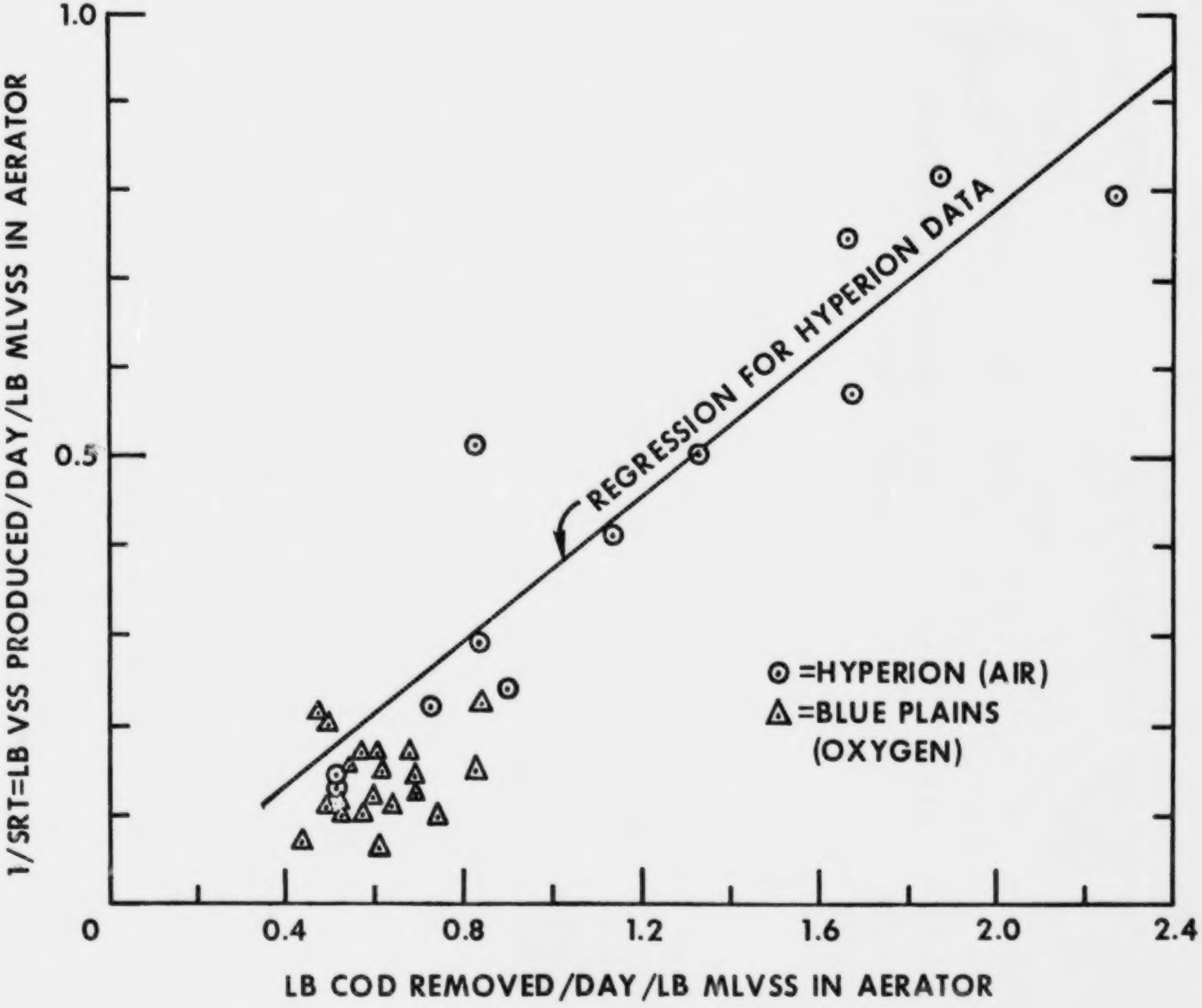


FIGURE 17. COMPARISON OF SLUDGE PRODUCTION FOR AIR AND OXYGEN SYSTEMS ON PRIMARY EFFLUENT FEED (COD REMOVED BASIS)

that reduced sludge production is probable with oxygen, but not to the degree predicted from the Batavia projects. Additional comparative data under similar operating conditions and on the same wastewater feed are needed. The current Detroit expansion is a good candidate for supplying such information when large-scale air and oxygen modules now in the final stages of construction and/or startup are ready for parallel operation.

Available sludge production data for Newtown Creek, Batavia, Las Virgenes, and the FMC project and two typical months of data for Blue Plains are summarized in three forms in Table 23 along with the corresponding F/M loadings. This table illustrates the general futility of attempting to correlate sludge production data from one location to another by any of the three methods shown. Many local factors including wastewater composition, wastewater temperature, mixed liquor D.O., and the biodegradability rate of various organic constituents combine to influence the amount of excess sludge that will be formed at different plants under similar loading conditions. Table 23 also shows that even at a single plant, unit sludge production on a BOD_5 or COD removed basis does not necessarily increase with increasing F/M loading, or vice versa. If any rational method exists for representing sludge production at a single location or comparing sludge production among locations, it is probably the basic inverse SRT method utilized in Figures 13, 16, and 17, or one of several published modifications of this basic method.

Power Consumption

One of the more attractive economic aspects projected for oxygenation systems from the Batavia work is greatly reduced power requirements for oxygen supply (generation) and transfer (dissolution) compared with the power requirements of air blowers. Estimated installed HP load requirements as projected from the Batavia work are plotted against plant size and BOD_5 aerator loading for both oxygen and air systems in Figure 18. The oxygen system band represents the additional power requirement for mixing as aerator volume is increased from one hour detention to two hours detention (based on Q). The air system band encompasses blower supply rates from 0.8 to 2.4 cubic feet of air supplied per gallon of wastewater

TABLE 23. SUMMARY OF OXYGEN SYSTEM SLUDGE PRODUCTION

Plant	F/M Loading $\left(\frac{\text{lb BOD}_5\text{/day}}{\text{lb MLVSS}}\right)$	$\frac{\text{lb Waste Sludge TSS}^a}{\text{mil gal}}$	$\frac{\text{lb VSS Produced}^b}{\text{lb BOD}_5\text{ Removed}}$	$\frac{\text{lb VSS Produced}^b}{\text{lb COD Removed}}$
Newtown Creek[c]				
Phse 4	0.92	1,300	0.94	0.49
Phase 5	1.19	1,185	0.73	0.59
Phase 6	1.62	1,055	0.59	0.51
Phase 7	2.44	1,025	0.60	0.46
Batavia[c]				
7/21/69-9/7/69	0.79	970	0.41	0.35
8/30/70-10/25/70	0.52	1,250	0.52	0.29
Las Virgenes[d]				
Phase 1	0.07	-	0.19	0.13
Phase 2	0.11	-	0.14	0.09
Phase 3	0.15	91	0.15	0.08
Phase 4	0.24	103	0.14	0.05
Phase 5	0.33	250	0.27	0.13
Phase 6	0.41	250	0.31	0.15
Phase 7	0.46	123	0.21	0.09
FMC Project[d]				
Phase 4	0.69	1,086	0.80	0.47
Phase 5	1.17	730	0.62	0.36
Phase 6	1.01	752	0.50	0.41
Phase 7	0.91	878	0.53	0.39
Blue Plains[d]				
Sept. 1971	0.39	620	0.38	0.25
Feb. 1972	0.30	430	0.47	0.23

[a] Includes waste sludge TSS only.

[b] Includes waste sludge and final effluent VSS.

[c] Raw sewage feed.

[d] Primary effluent feed.

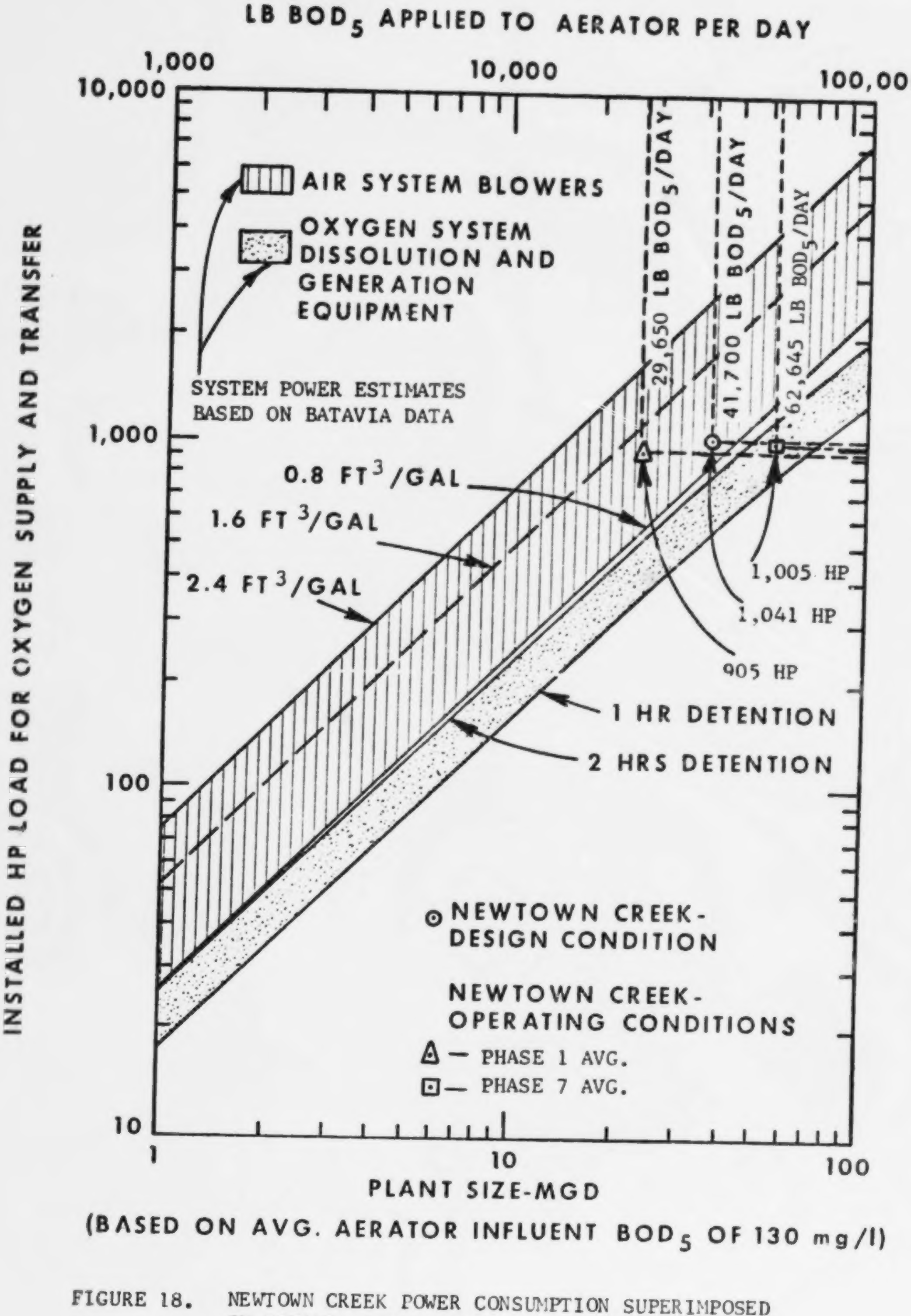


FIGURE 18. NEWTOWN CREEK POWER CONSUMPTION SUPERIMPOSED ON BATAVIA POWER PROJECTION CURVES

treated. Using the median of the bands, 50% and 65% reductions in aeration power requirements are projected for oxygen over air at plant sizes of 1 and 100 mgd, respectively.

Plotting the installed HP load for oxygen supply and dissolution at Newtown Creek on this curve indicates oxygen system power requirements estimated from Batavia may be somewhat optimistic. The installed HP load at Newtown Creek is 1041 HP, as broken down in Table 24, for a design load of 41,700 pounds of BOD_5 per day (calculated using a BOD_5 of 250 mg/l at 20 mgd). The Batavia curve predicts an installed nameplate requirement of only 550-800 HP for the same design BOD_5 load.

TABLE 24. INSTALLED HP AT NEWTOWN CREEK FOR OXYGEN GENERATION AND DISSOLUTION

Item	Nameplate HP
1. PSA Compressor	450
2. Liquid Oxygen Vaporizer	96
3. 1st Stage Compressors at 40 HP ea	80
4. 2nd, 3rd, and 4th Stage Compressors at 40 HP ea	120
5. 1st Stage Mixers at 50 HP ea	100
6. 2nd, 3rd, and 4th Stage Mixers at 30 HP ea	180
7. PSA Cooling Tower Pumps	8
8. Instrument Air Compressor	7
Total	1,041

The strength of the Newtown Creek wastewater during the project was somewhat weaker than the design projection. The actual BOD_5 load for Phase 1 averaged only 29,650 pounds per day (29 percent less than the design load) even though the average flow of 20.8 mgd was slightly higher than the design flow of 20 mgd. However, actual power consumption for the same period averaged 905 HP, a decrease of only 13 percent from the installed load. This demonstrates a well-known fact that non-variable speed drives operating below design load will consume almost as much power as when operating at design conditions. Consequently, power consumption for Phase 1 when superimposed on Figure 18 falls well above the oxygen band and up near the median of the air band.

The converse situation is illustrated in Phase 7 when, due to the increased average influent flow of 35.4 mgd, the BOD_5 load to the aerator averaged 62,645 pounds per day. Oxygen supply and dissolution requirements consumed an average power draw of 750 kilowatts (equivalent to 1,005 HP). Thus, with an actual power consumption 3.5 percent less the installed load, the system satisfactorily treated a BOD_5 loading 50 percent greater than the design load. For this phase, power consumption plots near the median of the Batavia oxygen band in Figure 18.

From the above data it is apparent that actual unit power consumption for oxygenation systems will approach installed unit power consumption only when operating at or near design organic load. Another conclusion which can be drawn from Newtown Creek experiences is that the oxygen module's oxygen transfer equipment was substantially overdesigned for the projected BOD_5 load.

Biological Performance Versus Biomass Loading

A strong point of oxygenation system performance noted wherever oxygen has been tested is the relative insensitiveness of effluent quality to changes in F/M loading. Data accumulated from Batavia, Newtown Creek, Blue Plains, and Las Virgenes are plotted in Figure 19. These data indicate plateaus for soluble BOD_5 and soluble COD breakthrough of only about 15 and 60 mg/l, respectively, up to F/M loadings of 2.4 lb total BOD5 applied/day/lb MLVSS. This denotes consistent and impressive performance under stressed conditions. At the lower loadings employed at Blue Plains, essentially complete insolubilization of BOD_5 is evident. For all F/M loading rates, however, total system efficiency for oxygen processes will be more directly dependent on solids capture efficiency (clarifier performance) than on biological performance deterioration.

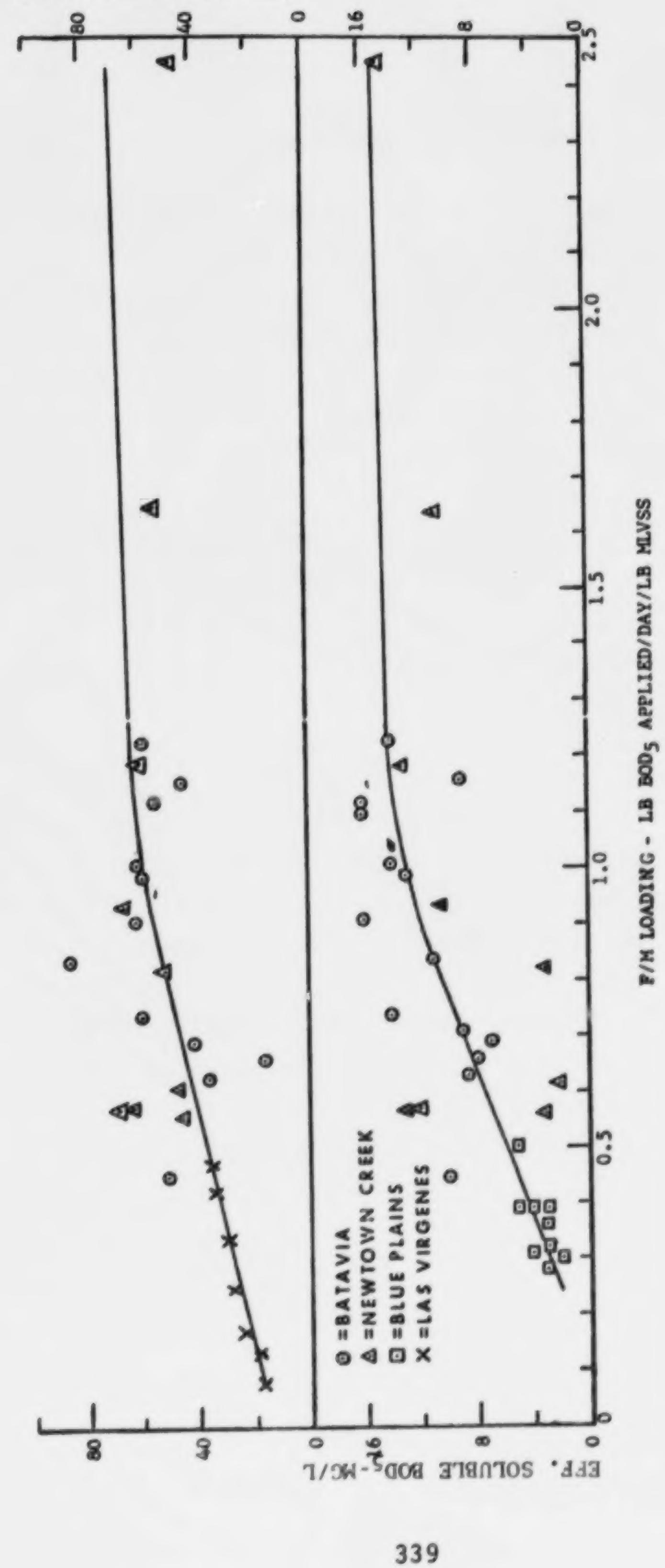


FIGURE 19. EFFECT OF F/M LOADING ON OXYGEN SYSTEM EFFLUENT SOLUBLE BOD_5 AND COD

SLUDGE SETTLING AND SYSTEM DESIGN

Perhaps the most important information generated by the Blue Plains project has been a delineation of some of the factors affecting sludge settling at that site and its resultant effect on system design (Stamberg, et al., 1972). In addition to the retardent effect on sludge settling previously mentioned due to filamentous infection of mixed liquor, other factors which affected oxygen sludge settling rates at Blue Plains included solids concentration, bulk sludge density (volatile solids fraction), and wastewater temperature.

In the range of MLSS concentrations at which hindered or zone settling occurs, it has been found that an equation in the form of $v_i = aC_i^{-n}$

where v_i = initial settling velocity

C_i = initial solids concentration

a = intercept constant

n = slope constant

when plotted on log-log paper yields a straight line. Further, it has been shown such a relationship exists for each of the three types of settling, discrete particle, hindered, and consolidation settling (Dick, 1970)(Duncan, et al., 1968). The change in slope between discreet particle settling and hindered settling normally occurs at a C_i between 2000 and 3000 mg/l. The hindered settling zone is characterized by a discrete subsiding interface and a zone of homogenously mixed settling particles. Clarifiers operating with initial hindered settling are in reality operating as sludge thickeners. It is essential that both hydraulic and mass loadings be considered in the design of secondary clarifiers for high solids systems. In many cases, thickening (mass loading) requirements will control the design. The best available approach for evaluating thickening requirements appears to be the batch flux (mass x settling velocity) method (Dick, 1970).

Bulk sludge density is a function of volatile solids fraction, i.e., density increases with decreasing volatile fraction. The incorporation of denser inerts into the sludge mass is the primary reason why biomasses developed on raw wastewater will generally settle better than those developed on primary effluent. Another manner in which sludge density is temporarily affected is the washing of accumulated inerts into a plant from its sewer system during rain storms. This point was vividly illustrated at Blue Plains during a tropical storm the summer of 1972 as shown in Figure 20 .

The least recognized parameter prior to plant startup that eventually strongly affected oxygen sludge settling rates at Blue Plains was wastewater temperature. Settling rates decreased significantly from summer to winter operation. For example, during September and October 1971 (a period when the oxygen clarifier was operated with a deep center feed below the sludge blanket to capture unsettleable particles) as wastewater temperature dropped from 81° to 71°F, the initial settling rate (ISR) decreased from 10 ft/hr to 7 ft/hr in a 1-liter graduated cylinder test at an MLSS concentration of 6000 mg/1 (see Figure 21). In November of the same year the center feed was raised above the blanket in an attempt to purge unsettleable particles from the system and increase bulk sludge density. While this technique did increase the sludge density and temporarily the ISR, a similar temperature effect was noted over the two-month period of November and December 1971. As wastewater temperature dropped from 70° to 63°F, the ISR decreased from 14 ft/hr to 9 ft/hr at an MLSS concentration of 4500 mg/1 (see Figure 22). Conversely, as Blue Plains wastewater temperature increased in the spring and summer of 1972, substantial increasing settling rates were observed as illustrated in Figure 23. The net result of this phenomenon was that a peak oxygen clarifier overflow rate of 1940 gpd/ft^2 was possible in the summer of 1970 with an MLSS of 8000 mg/1, while the peak overflow rate that could be

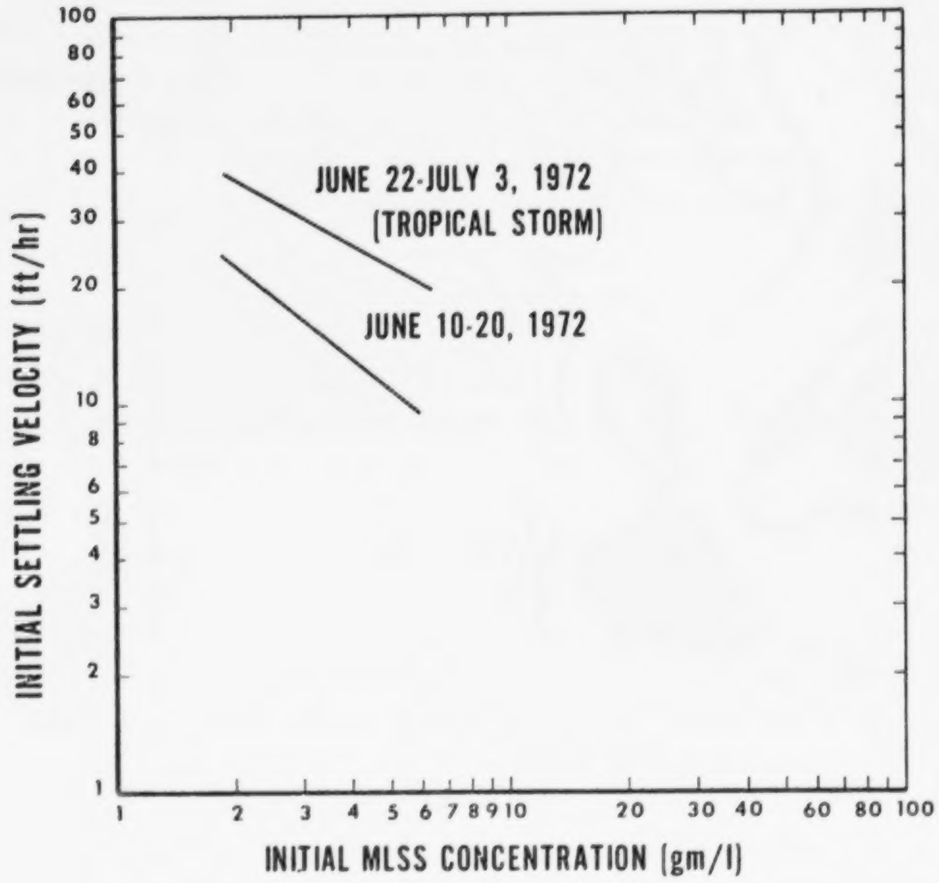


FIGURE 20. ILLUSTRATION OF INCREASED SLUDGE DENSITY CAUSED BY RAIN STORM AND ITS EFFECT ON INITIAL SLUDGE SETTLING VELOCITY[a]

[a] Reprinted with permission (Stamberg, et al., 1972)

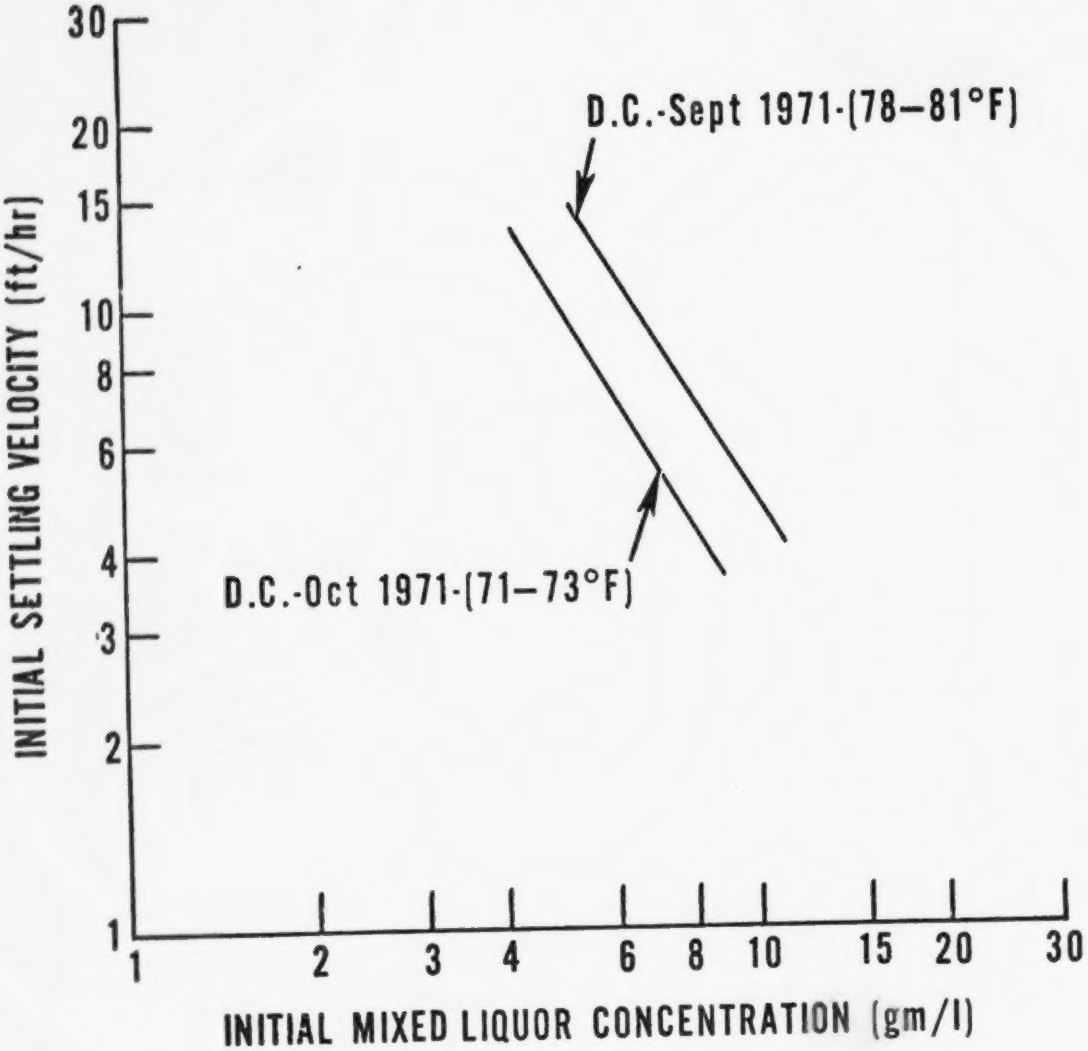


FIGURE 21. EFFECT OF DECREASING WASTEWATER TEMPERATURE ON INITIAL SLUDGE SETTLING VELOCITY (SEPTEMBER-OCTOBER, 1971)[a]

[a] Reprinted with permission (Stamberg, et al., 1972)

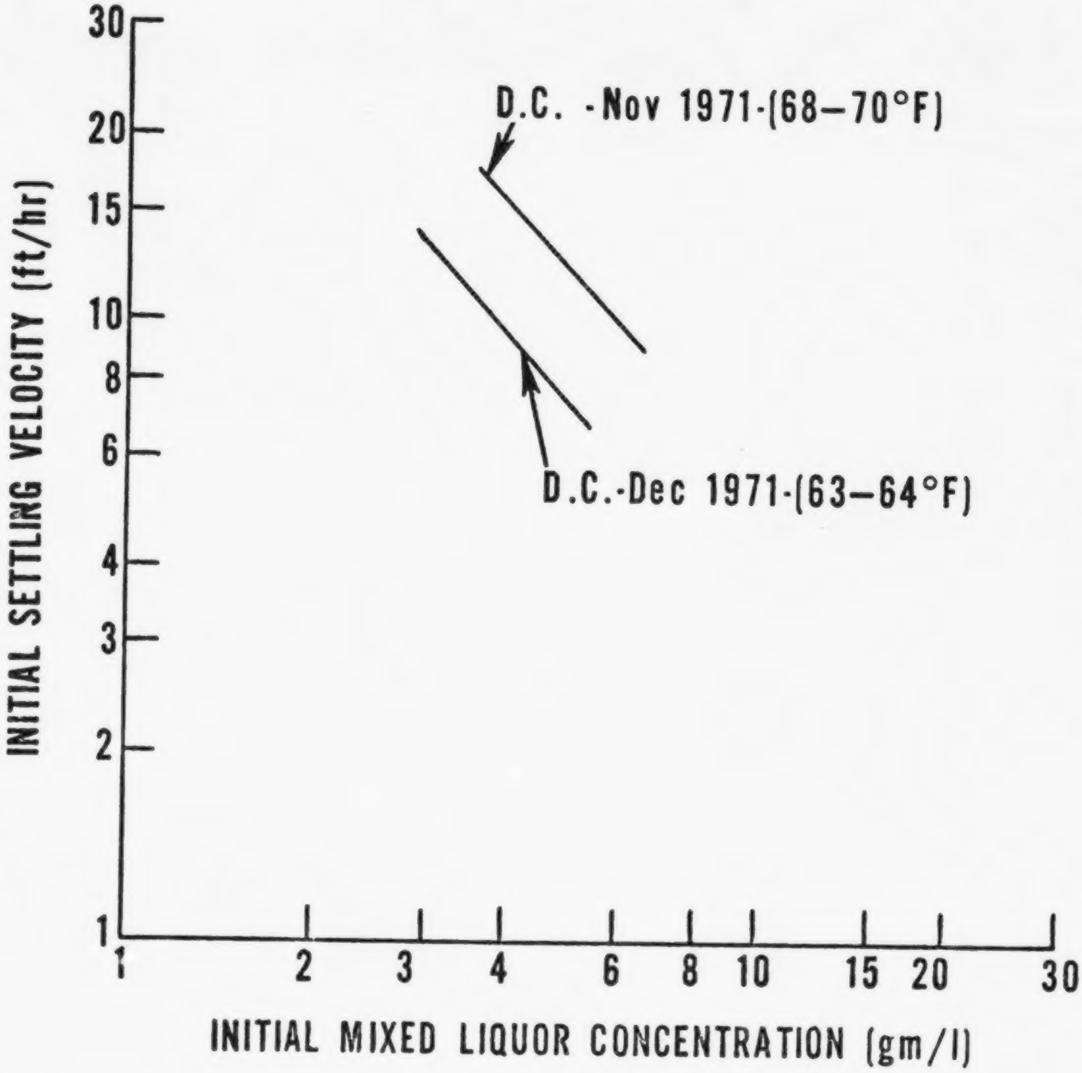


FIGURE 22. EFFECT OF DECREASING WASTEWATER TEMPERATURE ON INITIAL SLUDGE SETTLING VELOCITY (NOVEMBER-DECEMBER, 1971)[a]

[a] Reprinted with permission (Stamberg, et al., 1972)

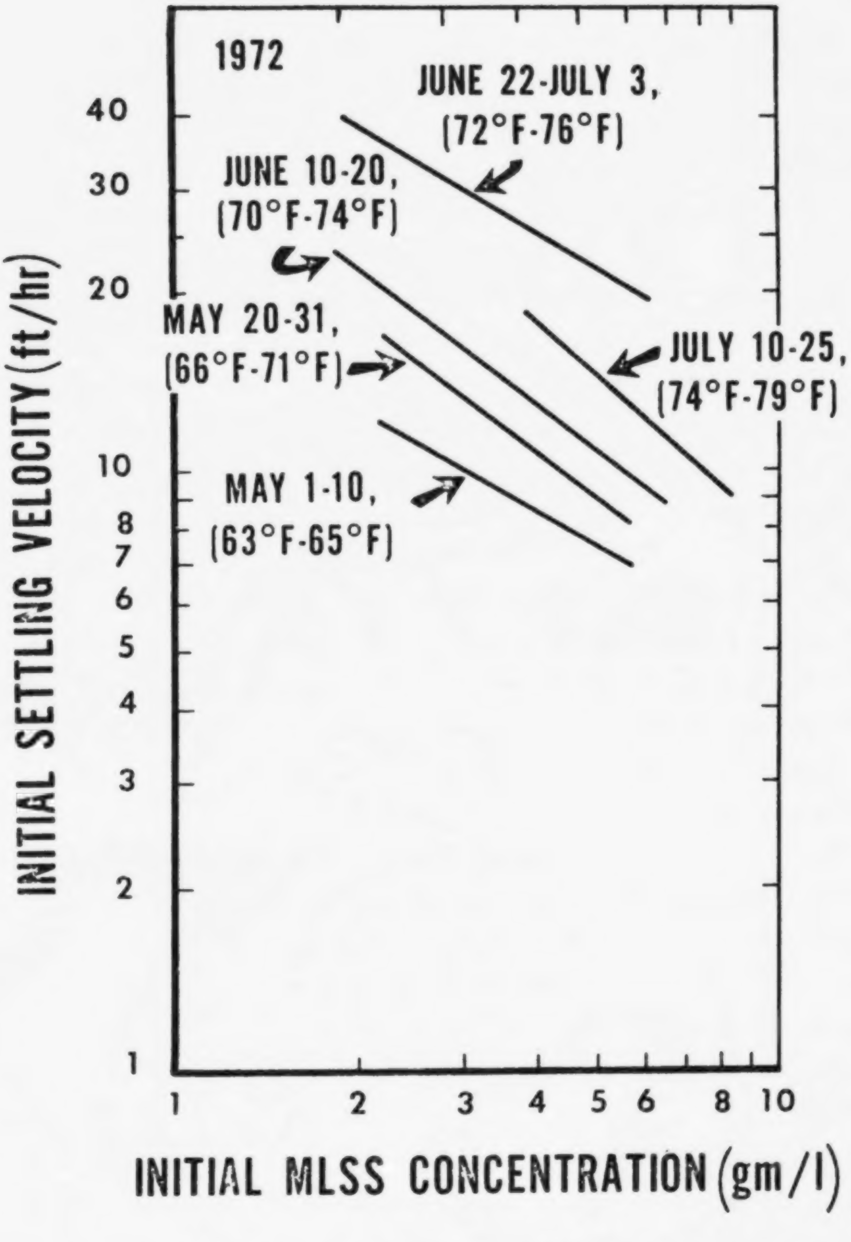


FIGURE 23. EFFECT OF RISING WASTEWATER TEMPERATURE ON INITIAL SLUDGE SETTLING VELOCITY (SPRING-SUMMER, 1972)[a]

[a] Reprinted with permission (Stamberg, et al., 1972)

sustained in either the 1970-71 or 1971-72 winters without clarifier failure was 975 gpd/ft^2 at MLSS levels that varied from 3900-5300 mg/l.

Undoubtedly all of the factors discussed above contributed to the reduced overflow rates necessary at Blue Plains to maintain satisfactory winter clarifier operation. However, it appears that wastewater temperature played a major role at this site. It is strongly emphasized though that the conclusions drawn from the Blue Plains project regarding wastewater temperature and sludge settling are not intended to imply that a similar effect will be noted universally. Much additional data are needed to reach a more definitive conclusion. Some additional data were collected over a period of about 20 months at Newtown Creek (raw wastewater feed) and Speedway, Indiana (primary effluent feed). Batch flux settling tests were conducted periodically at both sites using slowly stirred six-inch diameter, eight-foot long settling columns. Settling velocity profiles as a function of initial MLSS concentration are plotted in Fig. 24 for three runs conducted at Newtown Creek in December 1972 and June and August 1973. These plots tend to verify the temperature effect observed at Blue Plains. The decreased settling rates noted in the winter at Newtown Creek are probably due to a combination of increased viscosity and drag of the wastewater and alteration of biomass characteristics (proliferation of filamentous organisms) at the colder water temperature.

Results of the long-term Blue Plains work illustrate clearly that oxygen system design should be thought of as an integrated package consisting of a biological reactor, a secondary clarifier, and sludge handling facilities. The system should be designed for the worst anticipated climatic conditions at a given site. Clarifier sizing should be specifically tailored to the design and anticipated operating conditions of the reactor. There are two basic ways of achieving a desired F/M loading: (1) a small reactor and high MLSS or (2) a larger reactor and lower MLSS. If the first method is selected to save on reactor costs, a larger clarifier will be necessary. Both a small reactor and a small clarifier cannot be successfully mated in a design unless greatly reduced MLSS concentrations are utilized. However, opting for this selection will increase F/M loading, excess biological sludge production, and required sludge handling capacity and costs.

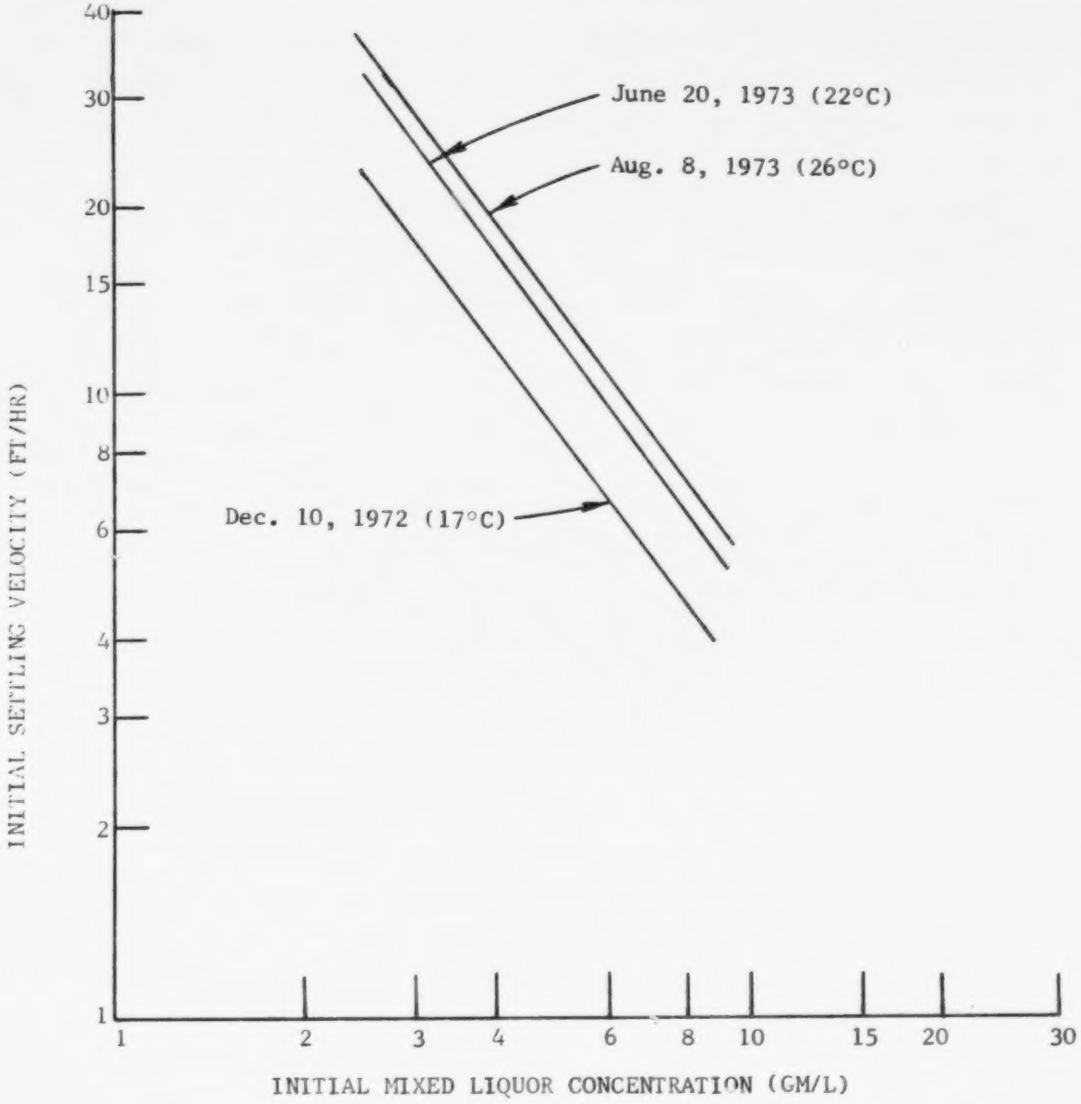


FIGURE 24. SETTLING VELOCITY PROFILES FOR BATCH FLUX SETTLING TESTS CONDUCTED AT NEWTOWN CREEK

PROCESS ECONOMICS

No updating of the comparative cost estimates presented for air and oxygen systems in the Batavia II final project report (17050 DNW 02/72) has been attempted in this paper. Figure 25 summarizes estimated total treatment costs (including amortization, operation, and maintenance) for air and oxygen systems of 1-100 mgd capacity as taken from the Batavia II final report. Interest was figured at 5-1/2% over 25 years. Oxygen supply costs are based on on-site generation plant purchase by the municipality. Projected savings in the cost of oxygen by buying and operating your own plant as opposed to commodity across-the-fence purchase current at the time of printing of the Batavia I final project report (17050 DNW 05/70) are shown in Figure 26.

Figure 25 projects average savings in total treatment costs of about 20% with oxygen for plants of 20-100 mgd. It must be remembered, however, that built into these curves are: (1) the assumption questioned by many observers that oxygen reactors will universally be one-third as large as air reactors for equal treatment and (2) what is believed to be overly optimistic estimates of the difference in sludge production rates between air and oxygen processes. The author concludes the one area in which oxygenation may have a very decided economic advantage is in the upgrading of existing overloaded secondary plants, such as Newtown Creek. Also it is likely that many decisions to install oxygen are not made so much on the basis of economics as on the basis of the high process reliability and stability and the rapid recovery following toxic upsets afforded by an enriched oxygen biological system.

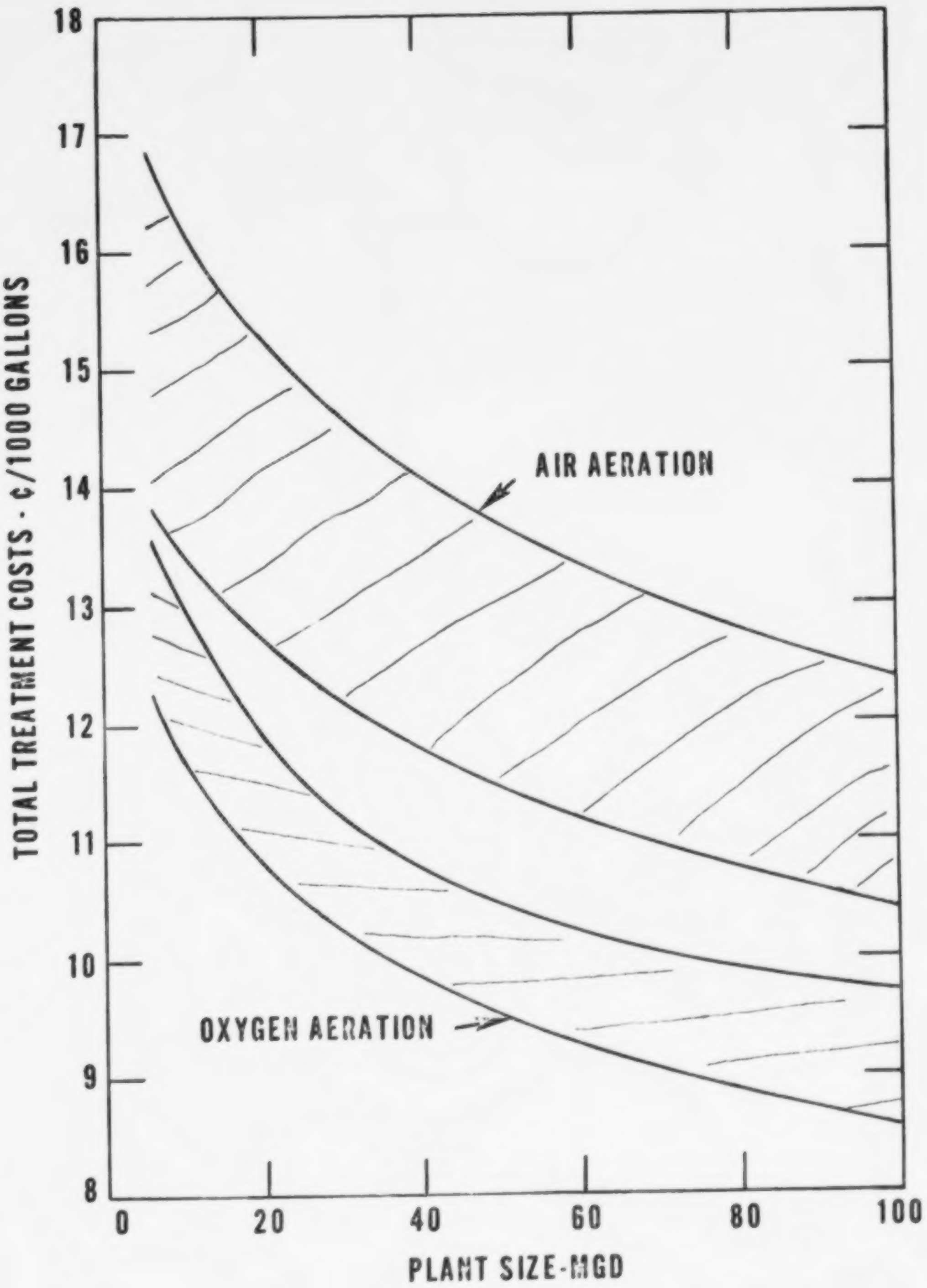


FIGURE 25. TYPICAL RANGES FOR TOTAL TREATMENT COSTS FOR NEW PLANTS WITH PRIMARY SEDIMENTATION PROJECTED FROM BATAVIA STUDIES

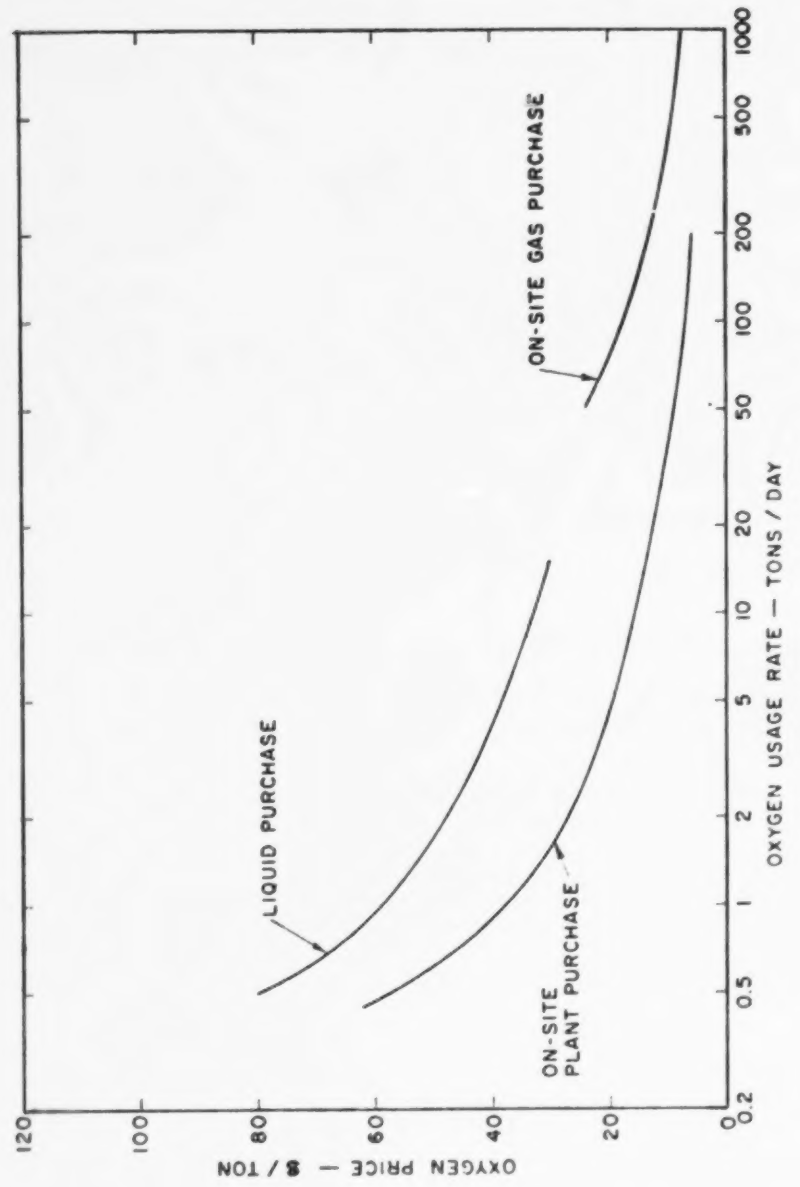


FIGURE 26. OXYGEN COSTS AS A FUNCTION OF USAGE RATE

CONTINUING DEVELOPMENTS

Continuing Research and Development Projects

Continuing EPA research and development efforts are underway or recently completed in the following areas: (1) evaluation of second-generation oxygen dissolution approaches, (2) examination of oxygen nitrification kinetics both in single-stage and two-stage systems, (3) definition of viable alternatives for combining chemical phosphorus removal with oxygen aeration, (4) determination of the most cost-effective sludge handling and dewatering techniques for taking advantage of the excellent thickening properties of oxygen sludge, (5) examination of sludge settling characteristics, (6) investigation of aerobic sludge digestion with oxygen gas, and (7) a study of the safety aspects of using oxygen in a wastewater treatment plant environment.

The development and maturation of new wastewater treatment processes are usually accelerated by the parallel development of several proprietary systems. However, because of this more rapid development, certain process details and aspects not directly associated with the treatment of wastewater often do not receive as thorough an evaluation as may be desirable and prudent. The safety aspects and requirements of utilizing oxygen in activated sludge treatment (No. 7 on the above list) are believed to represent one such aspect. Although each firm marketing an oxygen aeration system has undoubtedly considered safety features and requirements for its particular system, no comprehensive generalized treatment of the subject has been undertaken. until recently. Of particular concern is the processing of wastewaters which periodically contain hydrocarbons and other volatile substances in covered aeration systems with oxygen atmospheres ranging anywhere from 50 to 95%. The fundamental safety ramifications of using oxygen in this type of duty have needed an in-depth review and evaluation by an independent investigative team. A standard safety manual has also been urgently needed to instruct waste treatment plant designers and operators in the safe and proper handling of oxygen and to identify essential safety equipment and instrumentation. Such a manual must be sufficiently broad and comprehensive to apply to any rational concept for dissolving oxygen in wastewater.

Due to lack of funds, this project was delayed for more than a year. In June 1974, however, a contract was awarded to the Rocketdyne Division of Rockwell International to undertake this study. The purpose of the project is twofold:

(1) To evaluate the fundamental ramifications and implications from a safety standpoint of using oxygen gas or oxygen enriched air for aeration of activated sludge systems and based on this evaluation to recommend an implementive course of action which will ensure the safety and security of wastewater treatment plant personnel and facilities.

(2) To develop a standard safety manual and safety checklist for the safe and proper handling of oxygen in a wastewater treatment plant environment that will apply in principle to any rational oxygen dissolution concept.

Oxygen Process Implementation

Oxygenation systems are being designed and constructed for many treatment plants across the country to meet a variety of new plant construction and plant upgrading needs. At the time of this writing, 48 known municipal oxygen systems are in various stages of design, construction, startup, or operation. As summarized in order of decreasing size in Table 25, the total design flow of these 48 plants is 2,714.8 mgd ranging in capacity from 0.9 to 600 mgd. Of the 48 plants,37 are designed using surface aerators for oxygen dissolution,eight with submerged turbines, one with fine-bubble diffusers (open-tank), one (Las Virgenes) employed a converted air blower and air diffusers, and one is still undetermined.

Oxygenation is also beginning to make inroads into the industrial wastewater treatment picture. As indicated in Table 26, eight systems are currently being constructed and/or operated to treat a variety of industrial wastes. The total design flow of these eight systems is 62.9 mgd ranging in size from 1 to 25 mgd. Six oxygenation systems are now also on-line in Japan.

Preliminary oxygen designs are being prepared for 30-40 additional plants still in the negotiating phase. It appears that oxygen aeration is definitely here to stay.

TABLE 25. MUNICIPAL WASTEWATER TREATMENT PLANTS UTILIZING OXYGEN

Plant	Design Flow (mgd)	Type of Dissolution System	Status (Oct. 1974)
1. Detroit, Mich.	600	Submerged Turbines	Design
2. Detroit, Mich.	300	Submerged Turbines	Startup
3. Philadelphia, Pa. (Southwest)	210	Surface Aerators	Design
4. Philadelphia, Pa. (Northeast)	150	Surface Aerators	Design
5. New Orleans, La.	122	Surface Aerators	Const.
6. Middlesex County, N.J.	120	Submerged Turbines	Const.
7. Oakland, Calif. (East Bay M.U.D.)	120	Submerged Turbines	Const.
8. Dade County, Fla.	120	Surface Aerators	Design
9. Louisville, Ky.	105	Submerged Turbines	Const.
10. Wyandotte, Mich.	100	Submerged Turbines	Const.
11. Denver, Colo.	97	Surface Aerators	Const.
12. Baltimore, Md.	73	Surface Aerators	Design
13. Tampa, Fla.[a]	60	Surface Aerators	Design
14. Miami, Fla.	55	Surface Aerators	Design
15. Duluth, Minn.	42	Surface Aerators	Design
16. Hollywood, Fla.	36	Surface Aerators	Const.
17. Cedar Rapids, Iowa	32.9	Surface Aerators	Design
18. Harrisburg, Pa.	31	Surface Aerators	Design
19. Springfield, Mo.	30	Surface Aerators	Design
20. Salem, Ore.	26.5	Surface Aerators	Const.
21. Danville, Va.	24	Surface Aerators	Const.
22. Euclid, Ohio	22	Surface Aerators	Const.
23. Ft. Lauderdale, Fla.	22	Surface Aerators	Design
24. Littleton/Englewood, Colo.	20	Surface Aerators	Design
25. New York, N.Y. (Newtown Creek)	20	Submerged Turbines	Oper.
26. Decatur, Ill.	17.7	Surface Aerators	Const.

[a] Two-stage oxygenation system.

(CONTINUED)

(CONTINUED)

TABLE 25. MUNICIPAL WASTEWATER TREATMENT PLANTS UTILIZING OXYGEN

Plant	Design Flow (mgd)	Type of Dissolution System	Status (Oct. 1974)
27. Fayetteville, N.C.	16	Surface Aerators	Const.
28. Chicopee, Mass.	15	Surface Aerators	Design
29. New Rochelle, N.Y.	14	Submerged Turbines	Design
30. Muscatine, Iowa	13	Surface Aerators	Const.
31. Winnipeg, Manitoba	12	Surface Aerators	Oper.
32. Fairfax County, Va.	12	Surface Aerators	Oper.
33. Brunswick, Ga.	10	Surface Aerators	Oper.
34. Tauton, Mass.	8.4	Surface Aerators	Design
35. Morgantown, N.C.	8	Surface Aerators	Const.
36. Lebanon, Pa.	8	Surface Aerators	Design
37. Fairbanks, Alaska	8	Undetermined	Rebid
38. Speedway, Ind.	7.5	Surface Aerators	Oper.
39. Deer Park, Tex.	6	Surface Aerators	Oper.
40. Lewisville, Tex.	6	Surface Aerators	Const.
41. Mahoning County, Ohio	4	Surface Aerators	Const.
42. Jacksonville, Fla.	3.4	Surface Aerators	Startup
43. Calabasas, Calif. (Las Virgenes, M.W.D.)	1.8	Converted Air Diffusers	Oper. Discontinued
44. Littleton, Colo.	1.5	Surface Aerators	Oper.
45. Cincinnati, Ohio	1.2	Surface Aerators	Const.
46. Hamburg, N.Y.	1	Surface Aerators	Oper.
47. Minneapolis, Minn.	1	Fine-Bubble Diffusers (Open Tank)	Const.
48. Chaska, Minn.	0.9	Surface Aerators	Oper.
	2,714.8		

---

[b] Treatment of Zimpro supernatant.

TABLE 26. INDUSTRIAL WASTEWATER TREATMENT PLANTS UTILIZING OXYGEN

Plant	Design Flow (mgd)	Type of Dissolution System	Status (Oct. 1973)
1. Container Corp.[a] (Fernandina Beach, Fla.)	25	Surface Aerators	Const.
2. Chesapeake Corp.[a] (West Point, Va.)	16.3	Surface Aerators	Oper.
3. Gulf States Paper, Inc.[a] (Tuscaloosa, Ala.)	10	Surface Aerators	Startup
4. Union Carbide Corp.[b] (Sistersville, W. Va.)	4.3	Surface Aerators	Oper.
5. Union Carbide Corp.[b] (Taft, La.)	3.8	Submerged Turbines	Const.
6. American Cyanamid Co.[c] (Pearl River, N.Y.)	1.5	Surface Aerators	Oper.
7. Standard Brands[d] (Peeksville, N.Y.	1	Surface Aerators	Oper.
8. Hercules, Inc.[b] (Wilmington, N.C.	1	Surface Aerators	Oper.
	62.9		

[a] Pulp and Paper

[b] Petrochemical

[c] Pharmaceutical

[d] Distillery

SUMMARY

1. Oxygenation systems are equally applicable to new plant construction and upgrading of existing overloaded secondary treatment plants.
2. Oxygenation systems should be designed as integrated packages consisting of a biological reactor, a secondary clarifier, and sludge handling facilities.
3. There are genuine indications that reduced excess biological sludge production is possible with oxygenation; however, additional verifying data are needed.
4. Research and development will continue on many areas of the total oxygenation process to fully exploit the potential of the process.

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AERATION SYSTEMS FOR METRO CHICAGO

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THIRD U.S./JAPAN CONFERENCE ON SEWAGE TREATMENT TECHNOLOGY
TOKYO, JAPAN

FEBRUARY 1974

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ABSTRACT

Aeration Systems for Metro Chicago

The Metropolitan Sanitary District of Greater Chicago has experimented with aeration system design and operation since 1920. Initial efforts involved full scale field studies on activated sludge in advance of the design of the three major treatment plants at North Side, Calumet and West-Southwest. Both diffused air and mechanical aeration were extensivelv evaluated. Evaluations included location in tank, tank configuration, rate of oxygen transfer, diffuser and aeration characteristics, and, finally, large scale operating problems. The District's present aeration practice evolved from this experimental program and many of the practices developed are employed in plant designs throughout the country.

The flow diagram for a typical District facility would include an air intake, primary and secondary air filters, centrifugal or positive displacement blowers, air transport system and porous diffuser plates located in the base of each aeration tank to provide a spiral flow regime. Sizes and quantities are presented for the District's design and operating practices and reference is made to detailed studies which led to the present technology.

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Aeration Systems for Metro Chicago

Bart T. Lynam*

Introduction

The Metropolitan Sanitary District of Greater Chicago provides wastewater treatment, water pollution control and flood protection services to an area comprising 860 square miles, including the City of Chicago and approximately 120 suburbs. Approximately 5.5 million people reside within the service area, which includes an industrial community contributing a wastewater discharge equivalent to 4.5 million additional persons.

Three major wastewater treatment works handle the bulk of the District's wastewater flow. The West-Southwest, North Side and Calumet Plants provided secondary treatment to 875, 350, and 200 MGD respectively in 1972. Each of these plants employ activated sludge treatment with fine-bubble aeration. The District also operates four additional plants in the 1 to 6 MGD range. Two of the plants employ fine-bubble aeration for secondary treatment while the others incorporate mechanical aeration.

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It should also be noted that the District is anticipating the start-up of its 30 MGD Salt Creek Water Reclamation Plant in early 1975. This highly instrumented facility with computer assisted control will also provide fine-bubble aeration to a two-stage activated sludge process for nitrification.

Development of Aeration Technology at the District

Early Years

When the Metropolitan Sanitary District was founded in 1889 its primary function was to prevent the pollution of Lake Michigan and the city's water supply. The Chicago river (which originally carried sewage wastes into the lake) was reversed and substantial quantities of lake water (up to 10,000 cfs) were diverted into the Illinois River Basin for dilution. Design and operation criteria was based upon the need for natural stream assimilation capacity for a population equivalent to 3,000,000 anticipated around 1920. Around 1910 the District initiated treatment plant studies in anticipation of the eventual population growth.

Experimentation with Secondary Treatment

Recognizing that natural stream assimilation was a temporary measure for a service area undergoing accelerated growth the District incorporated facilities for experimentally testing secondary treatment in the designs for the Des Plaines River and the original Calumet sewage treatment plants. Both went into operation in 1922.

The Des Plaines River plant incorporated full-scale activated sludge treatment while the original Calumet plant provided Imhoff treatment with a portion of the flow passing through full scale experimental activated sludge units.

Studies performed during the late 1920's at these plants included plain aeration, tapered aeration and sludge reaeration. Diffused aeration, mechanical aeration and various combinations of the two were evaluated. By the beginning of the 1930's the effect of process design and operational parameters, such as tank depth, residence time, return sludge rates, air flow rates per unit volume, and both spiral flow and ridge-and-furrow flow, had been investigated.

First Major Plant Design

The research activity at Des Plaines River and the original Calumet Plants resulted in the 175 MGD North Side Activated Sludge Plant which was placed in operation late in 1928. The design consisted of 36 diffused air, spiral flow aeration tanks providing a 6.3 hour detention time with 20% return sludge.

Two parallel rows of air diffuser plates were set in concrete holders at a depth of 15 feet. These diffusers were fabricated with a permeability ranging between 11.7 and 18.5.

Permeability is essentially defined as the air discharge in cfm passing a one-foot-square plate under a pressure of two inches of water. It should be noted that 18.5 was the highest permeability manufactured at that time. A plate ratio (surface area of plates/surface area of aeration tank) of 9.6% was employed. Compressed air was supplied to the system by 7 Turbo-Blowers having a combined capacity of 250,000 cfm at 7.75 psig. Oil-coated impingement filters were used to filter the incoming air.

Diffuser Plate Studies

An ongoing battery of diffuser tests were initiated shortly after the North Side start-up to provide an economic design for the proposed addition at Calumet and the new Southwest plant. These studies by Beck (1) investigated the effects of permeability and plate thickness on air rates, bubble size, and pressure loss. Efforts were aimed at reducing diffuser plate pressure losses in order to keep the cost of compressing air as low as possible. It was soon recognized that uniform air distribution through the diffuser plates was needed to avoid operational problems. Later it was found that plates with higher permeability gave better uniformity with only a slight reduction in bubble area per unit of air.

The studies also disclosed that lower pressure losses and reduced plate clogging, resulting in longer life, could be attained by operating the system with lower individual plate air rates. Perhaps, the most significant result of these early diffuser tests by Beck was the formulation of an accurate method for rating diffuser plates.

Second Major Plant Design

Based on the success of the system at North Side, a new 136 MGD activated sludge plant was placed in service at Calumet in 1935. The design for Calumet was similar to the North Side design except that the specifications called for diffuser plates with a higher permeability range (36 to 44). Contrary to the North Side experience, operations at Calumet were seriously handicapped by diffuser clogging due to industrial wastes containing suspended and dissolved iron compounds. These materials precipitated in the aeration tanks resulting in ferric oxide which clogged the upper faces of the diffuser plates. The clogging persisted for years. Extensive tests were made on raw sewage composition, diffuser properties and cleaning techniques. The problem was eventually solved by prohibiting discharge of pickled liquor into the sewer system and by increasing the air discharge rate through the diffusers to keep the particulate matter in suspension.

Operational Life of Diffuser Plates

By 1940 the Activated Sludge process was in service at the West-Southwest Plant in addition to North Side and Calumet. While iron deposits were hindering the Calumet Plant, diffuser clogging problems developed at West-Southwest. As a result, air distribution studies were initiated (2) at all of the District's major plants to determine what steps could be taken to extend the operational life of the diffuser plates.

The original diffuser plates installed at West-Southwest were replaced with plates of 80 permeability in 1945. By 1947 the pressure losses had again become excessive. Tests revealed that the plates clogged primarily from dust in the air supply. The air supply was particularly dirty in the area of the West-Southwest Plant due to soot from the coal-fired boilers. Unfiltered air contained as much as 9 mg of particulate/1,000 cubic feet. Filtered air from the original impingement air filters contained 3.5 mg/1,000 cubic feet. Many different filters and secondary filters were tested in order to reduce the dust content.

It was decided to replace the old impingement filters with Electro-Matic Filters in which unfiltered air passes through an electrostatic field.

Charged dust particles are then attracted by oppositely charged plates covered with an oil film, where they are ultimately captured. These filters were successful in reducing the dust content of the air to 1 mg/1,000 cubic feet. Early operation, however, pointed out the need to temper outside air in colder weather to prevent icing on the filters. Furthermore, the initial operation indicated that the 1 mg/1,000 cubic feet was still excessive.

By 1948 tests concluded that precoated bag filters should be installed at Southwest to reduce the dust in the filtered air to less than 0.1 mg/1,000 cubic feet. Additionally, the bag filter was simple to operate and maintain, had a low pressure loss across the filter, and was not adversely affected by inclement weather conditions. As a result it was decided to employ the bag filters as the primary filtering system and to utilize the Electro-Matic filters as a secondary system.

Evolved Aeration System at West-Southwest

In the design for the third activated sludge battery at West-Southwest an economical air diffusion system had evolved consisting of bag filters and diffuser plates with an 80-permeability rating utilizing a comparatively low air rate per plate. The low plate air rates required additional plates and hence a higher first cost. Yet by comparison the modified operating procedure was cost effective based on reduced maintenance costs and longer plate life. An exhaustive economic comparison was made for various types of diffusers (plates, jets, and tubes). The findings resulted in a system consisting of 1 inch thick porous ceramic plates similar to the replacement plates added to batteries A and B in 1945. The major change coming out of this study was the placing of the diffuser plates normal to the walls with 20% to 30% additional plates in the influent half of the tanks. When this change was considered with the improved air filtering the District's engineers anticipated that the plates would have a useful life of 14 years. Battery C was put in operation in 1950 and the original plates are still operating satisfactory today.

The excellent operation of Battery C resulted in the design of this system becoming the standard for the activated sludge process within the District.

The ease of operation, minimal maintenance required, and excellent performance resulted in its adoption as the aeration system for large plants.

Recent Work with Aeration

In addition to the design of a fourth activated sludge battery at West-Southwest, the District has engaged in aeration studies and applications in the following areas: stream aeration, basin aeration, small plant design and modification, and the Salt Creek plant design.

In order to meet regulations set by the IPCB requiring that certain set dissolved oxygen concentrations be maintained in the waterways system, the District considered a variety of in-stream aeration designs. The objective was to artificially aerate the low points on the oxygen sag curve along the District's Waterway system. Mechanical, diffused air and pure oxygen aeration equipment was studied, sized and evaluated. The design life was estimated at ten years since it is anticipated that the proposed Chicago Underflow Plan and major plant expansion to tertiary treatment would render the project obsolete.

The Chicago Underflow Plan (which would capture and treat the stormwater run-off from Chicago by utilizing temporary underground storage in tunnels and surface storage in abandoned rock quarries with gradual pump back to the major plants for complete treatment during dry weather) together with the tertiary treatment of all wastewater from the District's service area would insure adequate D.O. levels in the waterways without the need for in-stream aeration. An economic comparison indicated that total costs per pound of oxygen transferred were lowest for floating mechanical aerators primarily because of a lower capital cost. Unfortunately the narrow width of the channel precluded the design because of its potential navigational hinderance. A scheme, whereby a portion of the stream would be withdrawn into a pure oxygen aeration system and then reintroduced to the stream was evaluated. Although the entire system could be placed on shore out of the way of navigation, the total cost (per pound of oxygen added) was slightly higher and was thus passed over. Diffused aeration with porous diffusers was ultimately selected because of its location in the channel floor, out of the way of barge traffic. Its total cost fell between that for mechanical aeration and pure oxygen aeration.

District experience with aeration basin design and operation has been almost exclusively with mechanical aeration. For the most part, these aerators have had low capital costs, been easy to install and operate and are best suited for the relatively short design life of aeration basins.

In recent years the District has extended service to several remote areas which were struggling to provide treatment for their wastewater. In the course of expansion, several small, overloaded plants were obtained by the District through annexation. It was decided to employ mechanical aeration at two facilities because of their relatively short life spans in view of existing phase out schedules. In both instances improved treatment of the existing trickling filter plants required conversion to the activated sludge process. The existing circular trickling filters made it more economical and easier (as well as quicker) to install surface mechanical aerators.

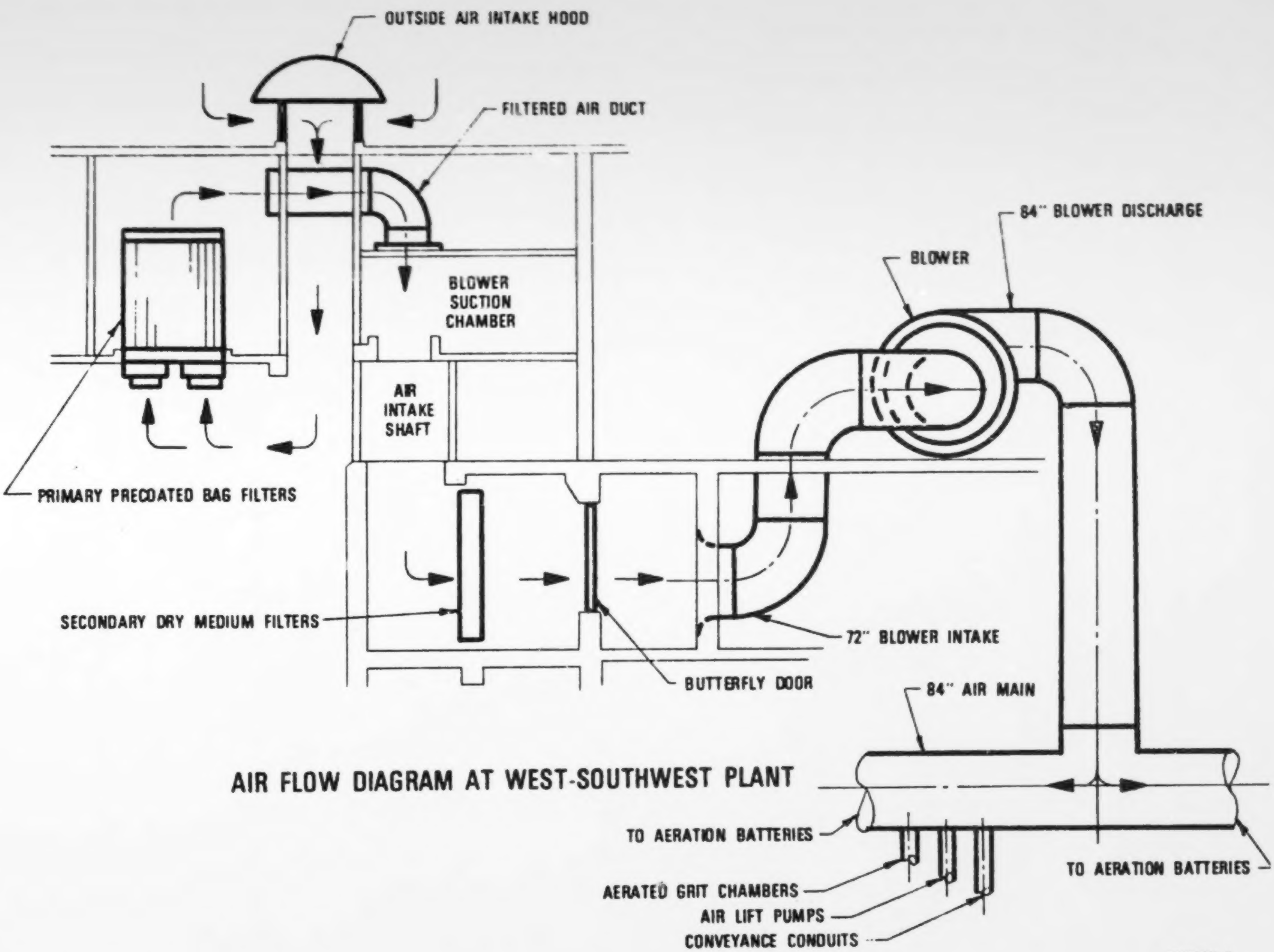
Most recently the District has undertaken the design and construction of a highly instrumented, two stage activated sludge Water Reclamation Plant at Salt Creek. Since the size (30, expansible to 125 MGD) was on the border line between a large and small plant, an economic comparison was made between mechanical and diffused air aeration. Mechanical aeration was slightly lower in total cost for the initial 30 MGD size. Both design concepts were then carefully reviewed. Ultimately, diffused air was selected because of (1) its economy in the light of anticipated plant expansion; (2) its capability to provide step feed, contact stabilization, complete mix and tapered aeration, in addition to conventional activated sludge treatment; and (3) its excellent past performance and reliability. This facility featuring computer assisted operation is scheduled to go on stream in early 1975.

FINE BUBBLE AERATION PRACTICE

Even though the equipment varies from plant to plant aeration systems throughout the District are conceptually similar. Although there are several uses for low pressure air within the plant, the primary use of the air is to transfer oxygen to the activated sludge process and keep the mixed liquor in suspension. In a typical District design, Figure 1, atmospheric air enters the system through an air intake. Air passes from the intake through primary and secondary filters before being compressed by the blowers. It then discharges into air mains and headers which transport the pressurized air to porous plate diffusers in the aeration tanks.

Intakes

Air intakes are located to optimize the air quality being drawn in and are usually equipped with louvers to protect the air passages. Atmospheric air passes through the intake into a chamber leading to the first of a dual set of filters. Besides connecting the intake and the first set of filters, the chamber also serves to blend warm building exhaust air with atmospheric air in winter to maintain a temperature above freezing through the filters. Without such provision condensate could freeze in the filters.



AIR FLOW DIAGRAM AT WEST-SOUTHWEST PLANT

FIGURE 1

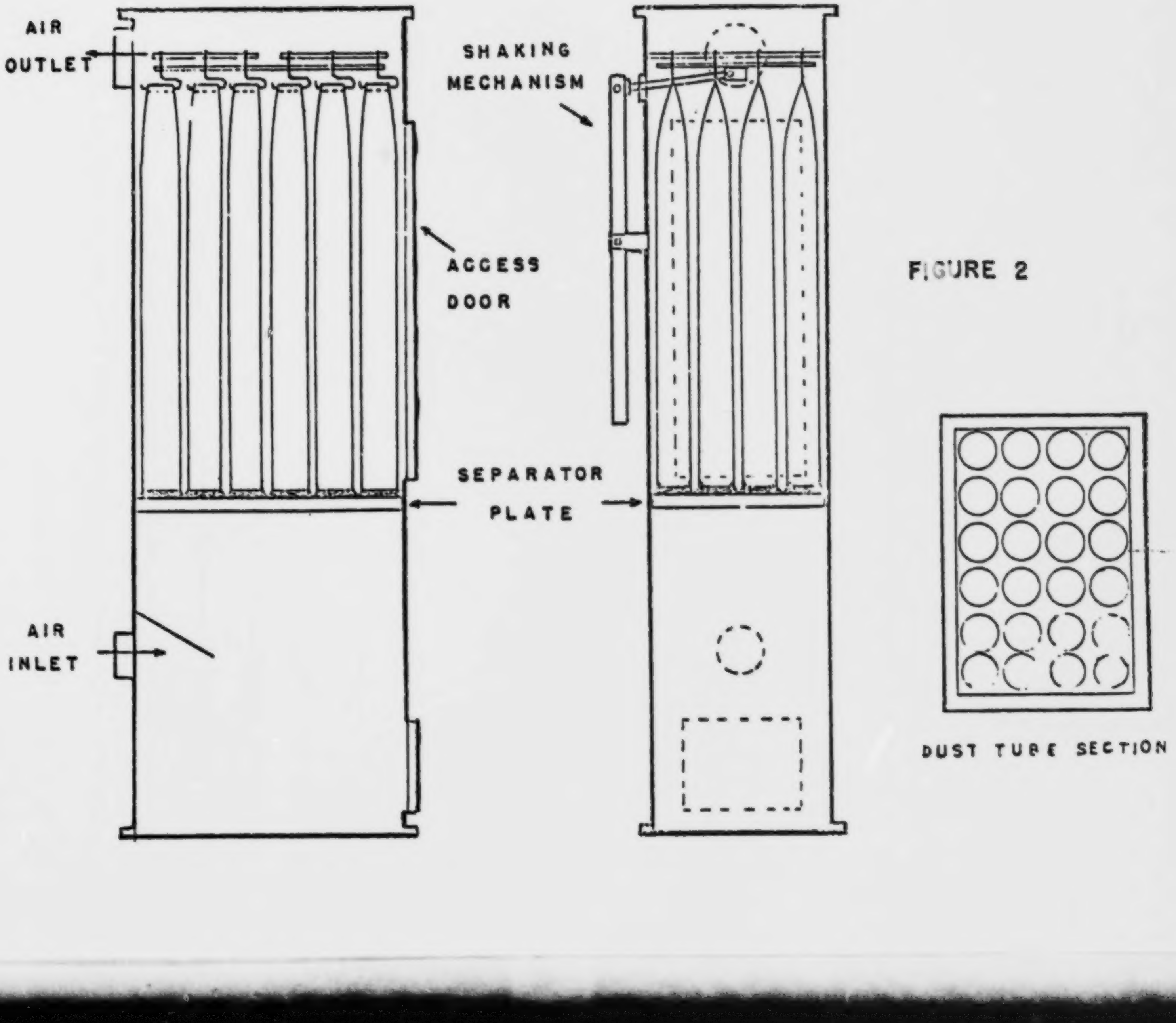
Filters

Air filters are the most vulnerable link in the aeration system's chain of processes. Basically, there are four types of filters which have been utilized over the years. These include viscous filters, dry medium filters, electronic filters and precoated bag filters.

The Metropolitan Sanitary District plants, like most plants practicing diffused aeration, employ a dual set of air filters. Primary filters remove most of the particulate matter. Secondary or back-up filters provided additional removal of particulates which pass the primary and protect against system damage, should the primary filters rupture or fail.

The primary filters at the West-Southwest plant are the precoated bag type equipment. Incoming air passes up through the center of the bag and then outward through the precoated material on the primary filter. During normal operation air passing the primary filters will contain particulate matter in concentrations less than 0.05 mg per 1,000 cubic feet. To insure good filtration the precoated bags are cleaned every year. One filter at a time is isolated and then shaken rapidly, Figure 2, so as to remove the precoat together with all accumulated dust.

FIGURE 2



DUST TUBE SECTION

The used precoat and dust falls into a hopper at the bottom of the filter unit and is removed. The bags are then recharged and the unit is returned to service. It is important to note that the absolute concentration leaving the filter is far more important than the percent removal by the filter. Filtered air quality is determined by drawing 50,000 cubic feet of filtered air through a tared sample thimble.

The secondary filters at the West-Southwest plant are of the dry medium type and have the shape of a large sock or mitten. These provide back-up and help to insure a long service life before cleaning and rejuvenation of the aeration system is needed. Furthermore filter equipment is over-designed in comparison to the blowers, since they comprise the critical stage in an aeration system; they are designed so that all blowers can be placed in service with one filter unit out of service. An important part of filter operation is a careful monitoring program. Important parameters include time in service, pressure drop across the filters, air volume filtered, air flow rate, and particulate concentration of the effluent.

Blowers

Filtered air makes its way through butterfly doors into a common air tunnel which feeds the plant blowers.

The District employs several types of blowers including vaneaxial, centrifugal and positive displacement varieties. The new Salt Creek Water Reclamation Plant will have a new type of blower which will permit a broader range of operating capability because of its automatic inlet guide vane control. The District's blowers range in size from 1,500 cubic feet per minute up to 200,000 cubic feet per minute. These blowers are powered by several types of prime movers including steam turbines, electric motors and gas turbines. Blowers, which are essentially low pressure compressors, produce an output pressure between 7.5 and 10 psi. Since the friction losses are in the order of 0.5 to 2.5 psi most of the blower output pressure is needed to overcome the 15 to 17 feet of hydrostatic head. Inherent in blower operation is a substantial temperature increase in the air. For example, with a blower discharge pressure of 8 psi the air temperature rise would be approximately 100° F above ambient temperature.

An important factor in the design and installation of blower systems is noise reduction. Silencers are installed in the air piping to maintain sound levels within a reasonable range.

In order to minimize noise on the largest blower at the West-Southwest plant, an acoustical housing was fabricated to enclose both the turbine and the blower. The insulation consisted of three (3) inches of fiberglass material.

Air Mains

The discharge from the blowers is transported through large air mains to the aeration tanks with small quantities being routed to other processes such as aerated grit chambers and air lift pumps. This air transport system consists of piping ranging from 84-inch conduits to 4-inch diffuser header pipes. Much of the piping is contained in service tunnels and galleries, thus enabling workmen to inspect and service them readily.

The general requirements for air piping are as follows: (a) the outside of the pipe must be sufficiently corrosion resistant for the location and (b) the inside of the pipe must remain so clean and sound that it will not give off any particles which would add to the clogging of the diffusers. Procedures have been developed for protecting air piping against corrosion and pitting. Large piping over 24 inches, which is located inside buildings, galleries, or embedded in concrete, is steel with the outside painted and inside bare.

The inside receives a grease coating during construction and is washed clean with solvents before being placed in operation. Smaller pipe, 8 to 24 inches, is spiral galvanized steel. Piping 6 inches and under is standard weight galvanized steel. Other piping which is submerged or immediately above wastewater is galvanized cast iron when 3 inches or larger.

These precautions are necessary for the following reasons. Air comes from the blowers at about 100° F above the atmospheric temperature, and as long as it remains a few degrees above the atmosphere there is no condensation to cause rusting. Therefore, no protective coating is necessary for mains inside the buildings and galleries after they are put in service. However, pipes too small for men to enter for thorough cleaning just prior to placing in service are galvanized to insure a clean interior. Bituminous coatings and paints are not used for interior coatings because oxidation of the coating or paint vehicle releases particles which contribute to diffuser clogging.

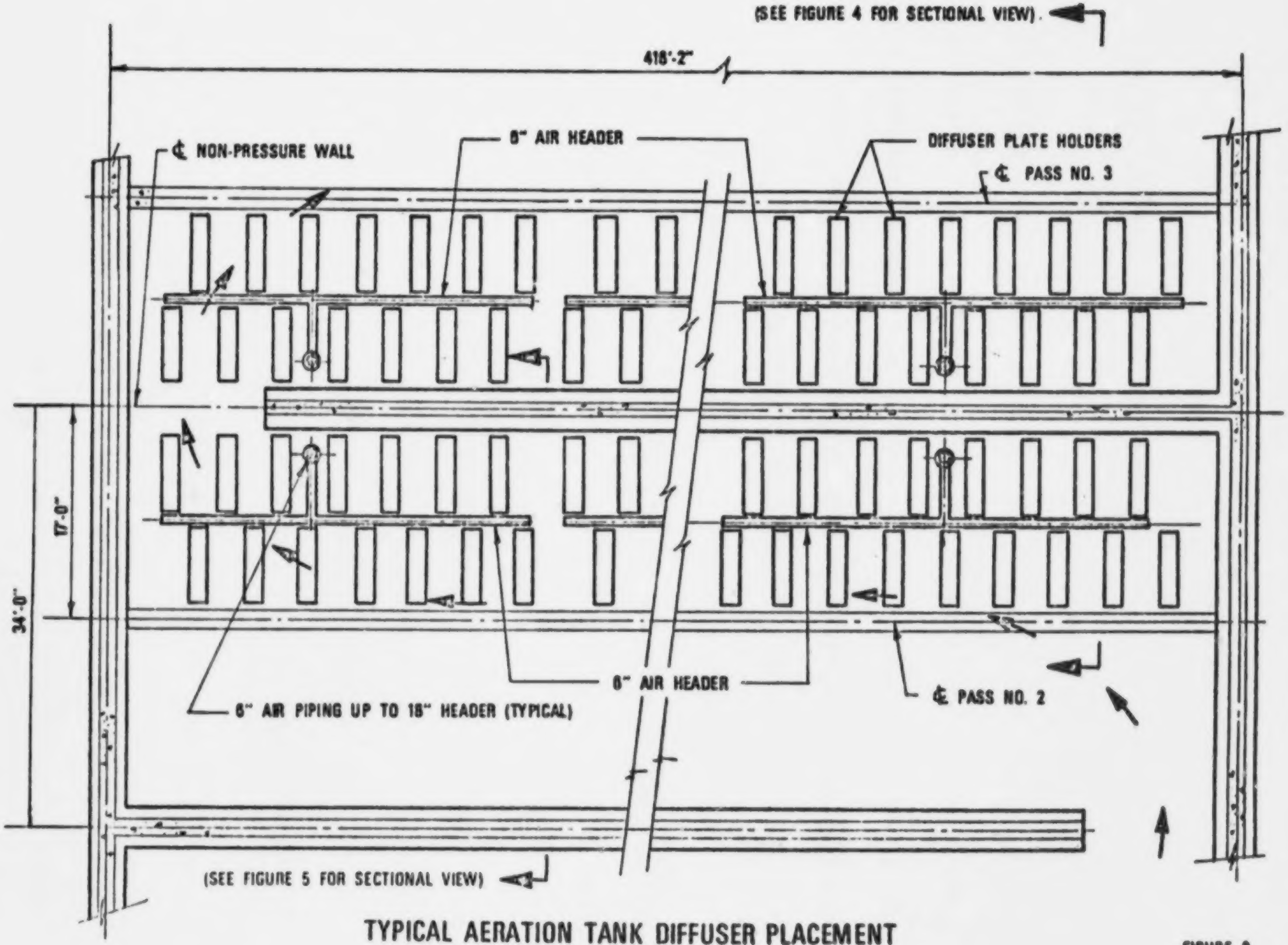
Where pipes are submerged, condensation will occur at times inside the pipes, due to the lowered temperatures, and this requires a good coating to prevent rusting.

Also, the outside of submerged pipes is subjected to corrosive conditions which are somewhat unpredictable.

Important factors in pipeline design are the types of couplings and joints. These must be capable of handling the velocities and temperature of the air flow. Extensive use of valving is incorporated in the system for both controlling and isolating air flow. Automatic and manually operated valves have been employed. The District's new Salt Creek Water Reclamation plant will have the capability of automatic air rate control. Dissolved oxygen sensors will regulate the automatic control valves which distribute air into the aeration tanks. Large diameter air mains at the District's major plants are usually fitted with butterfly valves, while the small diameter conduits consist of globe valves. An important function in operating the air transport system is the careful monitoring of air pressures and flows.

Aeration Tanks

Most of the aeration tanks, Figure 3, within the Metropolitan Sanitary District consist of long, multipass concrete channels having rectangular cross sections.

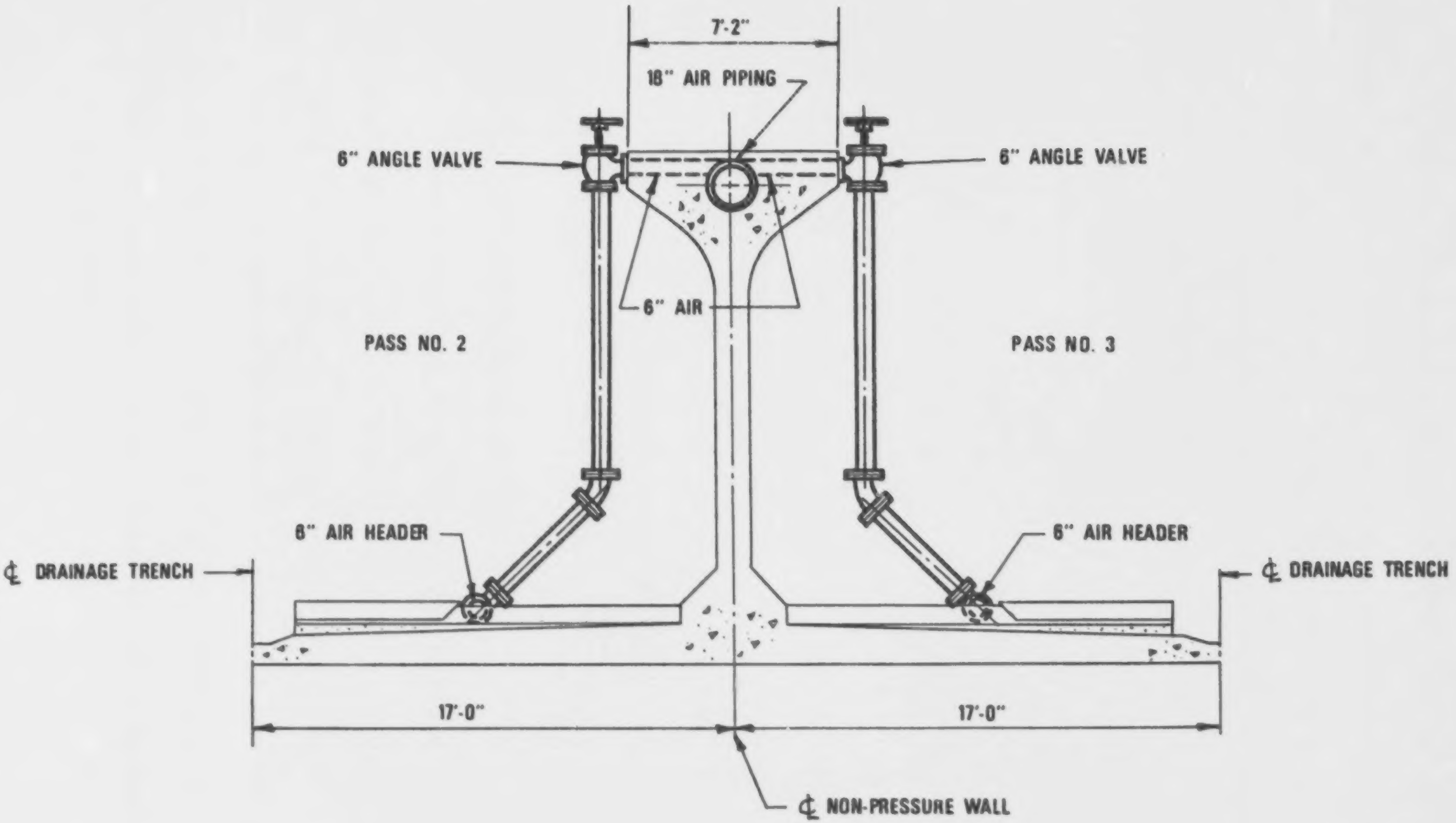


TYPICAL AERATION TANK DIFFUSER PLACEMENT

FIGURE 3

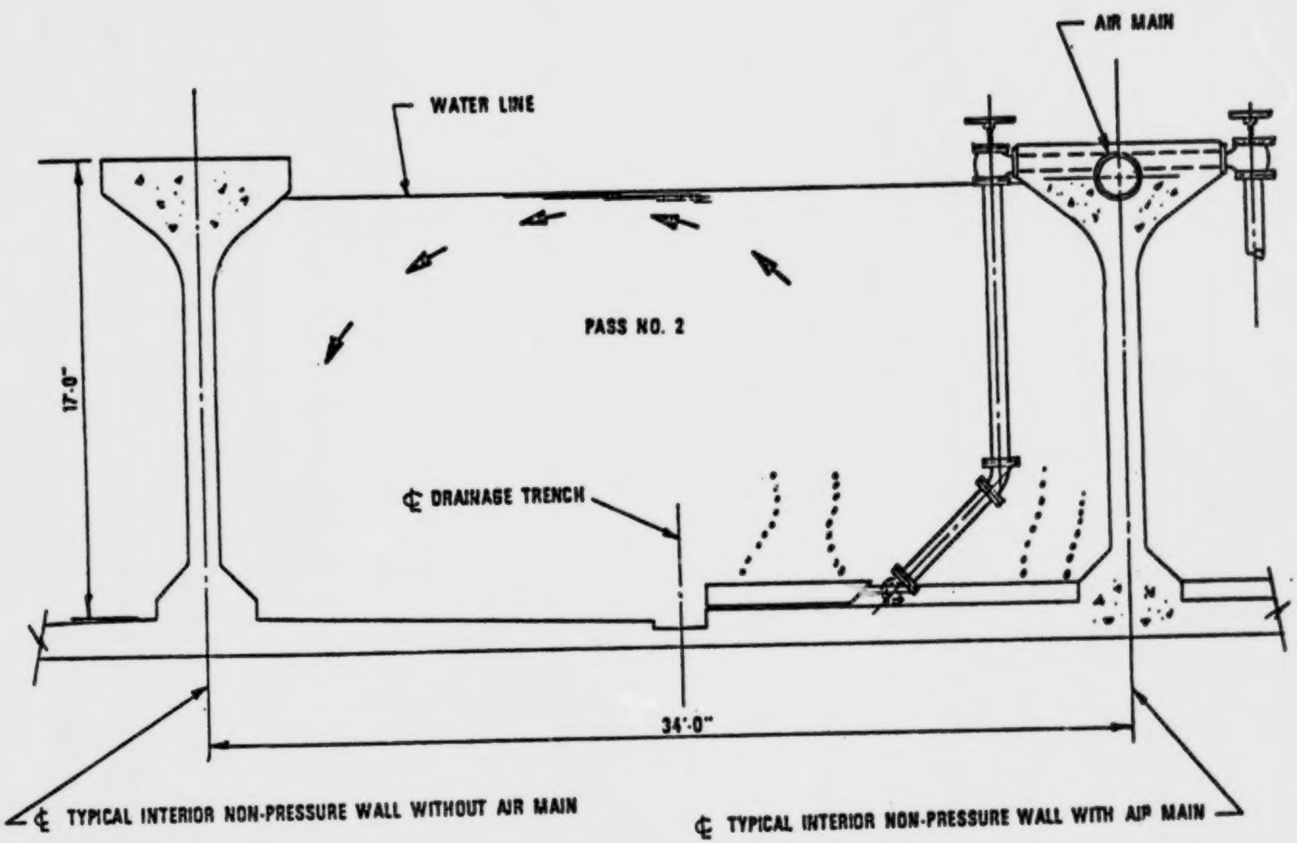
Porous plates tightly cemented in durable, solid concrete plate holders are placed (six plates/holder) along one of the channel walls. Because of its effective air filtration equipment, the District has been able to economize by setting diffuser plates with mortar rather than removable metal plate holders. A header pipe connects the air main and the porous plate diffusers, Figure 4, thus permitting distribution of air into the base of the aeration tanks. Since the diffusers are constructed adjacent to one of the channel walls, Figure 5, a spiral or overturning flow regime is established by the air distribution. As a result, two benefits accrue. One air main serves two channels and the spiral flow insures that the solids will remain in suspension. King (3) reported results of an extensive study of tank velocities at various horizontal and vertical locations.

Years of District experience with diffuser plates have led to well defined specifications for these diffuser plates. The plates are approximately 12" square and 1" thick. The fabrication materials may be either a crystal aluminum oxide bounded with a high alumina glass or a siliceous sand bonded with silicate. Rigid permeability (2) and oxygen absorption rating (3,4) requirements must be



TYPICAL AERATION TANK
CROSS SECTION

FIGURE 4


AIR MAIN
WATER LINE
PASS NO. 2
17'-0"
℄ DRAINAGE TRENCH
34'-0"
℄ TYPICAL INTERIOR NON-PRESSURE WALL WITHOUT AIR MAIN
℄ TYPICAL INTERIOR NON-PRESSURE WALL WITH AIR MAIN

TYPICAL CROSS SECTION OF SPIRAL FLOW AERATION TANK

FIGURE 5

met as determined by specific empirical techniques.

After installation, pressure loss and volume of air diffused are monitored closely. From time to time, the air rate is temporarily raised and lowered to enable a plot of pressure loss versus flow rate. A pronounced upward curve is an indication that the plates are clogging. Clogging can occur on the air side or the liquor side. Except for the ferric oxide deposits at the Calumet plant, most clogging has occurred on the air side. Besides particulate problems, some pressure loss has occurred from condensation in diffuser headers during the early spring and summer. Two methods have been employed for rejuvenating the plates. The most effective has been removal and kiln burning. The second, a temporary measure, is washing the plates with an acid.

Recent District designs provide more than adequate mixing velocities. This is insured by plate arrangements which discharge 8 to 12 cfm of air per foot of tank length. Several plates are placed side by side in rows to provide a band of aeration between 6 and 12 feet in width. In the past, fewer plates were placed in the effluent half of the aeration tanks to produce a tapered aeration effect.

This served well in the conventional activated sludge process. The District now provides step feed capabilities in its designs so a uniform plate area is provided throughout the aeration tanks. The nominal distribution capacity of an individual diffuser plate ranges from 1 to 4 cfm. MSD experience has demonstrated that a long service life can be obtained from the porous plates if adequate air filtration and careful placement of the plates are practiced.

Other Uses of Low Pressure Air

The aerated grit chambers in the District plants utilize air drawn from the blower system. This air is diffused into the wastewater stream through submerged non-porous diffusers located close to the bottom of the chamber. The quantity of air used by the aerated grit chambers ranges from 2½ to 5 percent of the total blower output. Another important use of low pressure air is the air lift pumps serving the final clarifiers, primary settling tanks and concentration tanks. The quantity utilized in these locations ranges from twelve to twenty percent of the total blower output.

Low pressure air is also used in minor quantities for aerating and mixing various wastewater conveyance conduits. Moreover, it is utilized for post aeration at the Hanover Park Water Reclamation Plant and the Salt Creek Water Reclamation Plant.

SUMMARY

The Metropolitan Sanitary District of Greater Chicago has been engaged in aeration systems development since 1920. Its principal contributions have been in diffuser permeability criteria, design and operation of primary and secondary air filters, aeration tank configuration and diffuser placement and the formulation of mathematical relationships for oxygen transfer. The result has been an effective economical technology for the activated sludge process.

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OXYGEN ACTIVATED SLUDGE SYSTEMS IN TEXAS

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OXYGEN ACTIVATED SLUDGE SYSTEMS IN TEXAS

REVIEW OF OXYGEN USE IN WASTEWATER TREATMENT

Since D. A. Okun's first experiments with the Bio-Precipitation, a pure oxygen process, in the late 1940's, the use of high purity oxygen in wastewater treatment has created surges of interest followed by periods of diminished interest in the academic community and in industry. These early studies were conducted on laboratory scale and on pilot plant scale during the 1950's. The pilot plant sutdies on municipal sewage concluded that the use of pure oxygen resulted in a savings in reactor volume and total plant area. These savings were due to the greater concentration of mixed-liquor solids that can be maintained aerobic with oxygen. The power requirements of the conventional system and the Bio-Precipitation process were found to be essentially equal. In order for the Bio-Precipitation process to offer an economic advantage, the cost savings resulting from the decreased size of the reactor had to be greater than the additional costs associated with providing high purity oxygen. Evidently, the advantages were not sufficient, as further research into the use of pure oxygen was not significant for nearly ten years.

Interest in the use of pure oxygen was revived by the Linde Division of Union Carbide Corporation in the late 1960's, when the Corporation began seeking additional markets for its new oxygen production process. The Union Carbide process utilized a multi-stage, covered system which it named the Unox Process. This system was first evaluated at the Batavia, New York, municipal sewage treatment plant where it was operated in parallel with a conventional activated sludge plant.(3) Since that initial investigation, the applicability of the process to a wide range of wastewater characteristics has been demonstrated by pilot plant studies throughout the United States.(1)

THE UNOX PROCESS

The only significant application of pure oxygen to waste treatment available today is the Unox Process. In the Unox Process, oxygen absorption takes place within the mixed-liquor of the biological reactor. The system utilizes a series of gas-liquid contacting units or cells. Mass transfer and mixing within the cells is accomplished with a surface aerator or a sparged-turbine system. The individual cells are gas-tight, allowing the atmosphere in each cell to be collected and diffused into each succeeding cell. High purity oxygen is added to the first cell, thus providing maximum dissolved oxygen concentrations in the first cell where the oxygen demand is greatest. At the end of the series of cells, the remaining gas, containing a small fraction of the initial oxygen content, is vented to the atmosphere. Liquid-solids separation is accomplished in conventional clarifiers.

THE UNOX PROCESS IN TEXAS

There are presently no large-scale Unox systems in operation or under design in Texas. Four small to medium size Unox systems are under construction or are being planned by Texas municipalities. A 6 MGD (22,800 m^3/day) Unox plant is being built by the City of Deer Park. This plant is scheduled to begin operation in the early part of 1974. Construction on another 6 MGD (22,800 m^3/day) plant for the City of Lewisville began in January 1974. Additional Unox installations are planned by Pasadena and Missouri City, Texas.

There is no plant performance data available for any of the Texas Unox installations at this time.

REPORTED ADVANTAGES OF UNOX

The reported advantages of the Unox Process include: reductions in biological reactor volumes, higher oxygen-transfer efficiencies, improved sludge handling characteristics, simplified process control and odor control.

Reduction in the size of the biological reactors is accomplished by the use of a much higher mixed-liquor solids concentration in the Unox Process. The higher concentration of biological solids can be maintained in an aerobic condition by the greatly improved oxygen mass-transfer efficiencies of pure oxygen. The higher mixed-liquor solids concentration allows the designer to decrease the biological reactor volume while maintaining the system mass-loading rate at a level comparable to that of conventional designs. The savings realized from tankage reductions represents the main economic advantage of the oxygen system.

The more direct advantages resulting from the higher oxygen-transfer efficiencies of pure oxygen include equipment-size reductions and reduction in power usage. The lower equipment requirements of the oxygen system produce a savings in a projects capital cost, and the reduced power consumption a decrease in operating costs.

Other reported advantages of oxygen are improved sludge settling characteristics and a reduction in the quantity of waste solids produced.

The greatly improved sludge settling characteristics of oxygenated biological solids have been reported by a number of investigators.(3) The advantages of an improved settling sludge are twofold. The improved clarifier-underflow concentration allows the maintenance of the high operating mixed-liquor solids concentrations and the thicker clarifier underflow results in considerable savings in sludge handling.

The Union Carbide Corporation has reported a reduced sludge production from their Unox Process and has attributed the decrease to a highly aerobic sludge. They have postulated that the high dissolved oxygen concentration of the Unox Process is able to penetrate large floc particles, thus maintaining the entire sludge mass aerobic and active.(1) These results have been contradicted by independent investigations conducted on bench-scale-sized pure oxygen units.(3) The results of these experiments indicate the Kinetics of activated sludge systems operated at a dissolved oxygen concentration as great as 20 mg/l do not vary significantly from the reported Kinetics of conventional systems. Furthermore, studies of oxygen diffusion through biological floc particles have shown that for dissolved oxygen concentrations and floc particle sizes normally encountered in the conventional activated sludge process, the entire mass of the floc particle is aerobic.(2)

The dispute over excess solids production should be conclusively answered in the next several years as an inventory of operating data is accumulated.

The simplified process control of the Unox system also has its advantages. A simple pressure control system is used to monitor the oxygen requirements of the system, thereby permitting only the oxygen needed to be applied. The system easily "tailors" the oxygen applied to the diurnal demands of the wastewater. A considerable savings in power is realized by this integrated system.

The enhanced ability of the Unox Process to respond quickly and automatically to varying oxygen demands, thereby maintaining a minimum dissolved oxygen concentration of 1-2 mg/l, and to maintain higher return sludge solids concentrations should improve the stability of the process and minimize plant upsets. Should this be the case in actual plant operations, the Unox Process would have a decided advantage over the conventional activated sludge process.

Another advantage of the covered tank system is that effective odor control of the vent gas can be practically achieved. The total gas vented from the system is reported to be about 1% of the gas vented from an air-activated-sludge system.(1)

POSSIBLE DISADVANTAGES OF THE UNOX SYSTEM

The Unox Process includes sophisticated mechanical and electrical control systems that ordinary treatment plant operators may find difficult to control and maintain. It is a sad fact, but true, that many of the treatment plant operators in the State of Texas and in other states are undereducated and underpaid. These personnel may not be capable of maintaining the Unox Process even after extensive instruction. Municipalities located in the industrial-urban areas of the State, where there is a reservoir of persons trained in the maintenance of instrumentation and complex machinery, can more readily assume this risk.

Another possible disadvantage of the Unox Process is the possibility of volatile hydrocarbons entering the enclosed system which could lead to a serious explosion. This possibility has been recognized by Union Carbide and a sensitive hydrocarbon monitoring device is installed in every Unox unit sold. When significant hydrocarbon concentrations are detected by the sensing device, the tanks are automatically purged of oxygen with air. Proper maintenance of this monitoring device would seem to be mandatory.

FUTURE OF UNOX IN TEXAS

The Unox Process seems to be a viable alternative to the conventional activated sludge process. Its greatest advantage seems to be the reduction in tank volumes and land areas needed for treatment. These advantages would be of primary importance in the renovation and expansion of existing large sewage treatment plants in densely populated areas.

While the economic advantages of the Unox Process for large municipal plants may be considerable, there may be no economic advantages for small communities. Small communities should carefully consider the economics of the Unox Process as compared to conventional designs, with special attention being given to the complexity of operation and the need for highly trained operators. In Texas, we are assuming a "wait and see" attitude before reaching a decision on the merits of this process in actual operation.

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SUSPENDED SOLIDS REMOVAL

PROCESSES STUDIED AT METRO CHICAGO

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SUSPENDED SOLIDS REMOVAL PROCESSES STUDIED AT METRO CHICAGO

The Metropolitan Sanitary District of Greater Chicago was created by the Illinois State Legislature in 1889, to protect Chicago's Lake Michigan drinking water supply from pollution. This action was deemed necessary because in 1885, a heavy storm of more than six inches over the Chicago area in a two day period, flushed the streets, catch basins and sewers into the rivers and, subsequently, polluted the lake far beyond the intake cribs which supplied the city's drinking water. As a result, approximately 12 percent of the city's population died from such waterborne diseases as cholera, typhoid and dysentery.

The initial treatment system consisted of reversing the flow of the Chicago River by the construction of a man-made canal system, so that it carried discharged polluted water away from the lake. This system of treatment by dilution was the first step in preventing raw sewage from entering Lake Michigan and providing a minimum level of treatment.

Although the reversal of the Chicago River solved the problem for Chicago, neighboring states on the great lakes protested that the District was draining the lake and sued the District to prevent it from taking unlimited quantities of water from the lake. As a result of the law suit, lake diversion was reduced from 10,000 to 3,000 cubic feet per second. Therefore, controlling locks were constructed at Wilmette Harbor, the mouth of the Chicago River and the Calumet River. These locks control the diversions from Lake Michigan.

The loss of diversion water and the constantly increasing flows of the rapidly growing Chicago area required the building of treatment plants to intercept and process the sewage flowing into the canals in order to keep the rivers and waterways from becoming overly polluted. Research work initiated by the District in 1911 enhanced by urging of the Supreme Court dealing with the diversion matter, commenced a second phase in the District's waste treatment program. The North Side plant, serving the northern portion of the area, was completed in 1928 and expanded in 1937. The West-Southwest plant, the world's largest sewage treatment plant, was placed in operation in 1939. Together with the Calumet plant these plants today treat over 1.3 billion gallons of sewage from a population of about 6.0 million people and an equivalent of an additional 4.5 million people from non domestic sources.

The District recognizes, however, that today secondary treatment cannot be considered to be the ultimate objective in sewage treatment as it was in the past. In addition, stricter rules and regulations being imposed by the Illinois Pollution Control Board (IPCB), makes it mandatory to discharge a higher quality effluent relatively free of contaminants including organic suspended solids.

IPCB rules and regulations take on added significance to the Sanitary district when one considers that, during periods of low flow up to 99% of the flow in the sanitary district's man-made controlled waterway system consists of MSDGC effluent.

According to the newly adopted rule, section 404 (f) (ii), MSDGC is subject to an effluent standard of 10 mg/l of BOD_5 and 12 mg/l SS by December 13, 1977. For these reason the District has been actively engaged in studying advanced wastewater treatment processes on pilot or full-scale operation since 1968.

Because of the importance of effective suspended solids removal to the production of high quality effluents, the MSDGC has conducted extensive full scale and pilot studies in this area. This paper presents some of the Districts experiences resulting from these studies.

FULL SCALE EVALUATION OF SAND FILTRATION AND MICROSTRAINING AT HANOVER PARK

INTRODUCTION

The initial full-scale tertiary treatment studies in terms of removal of solids were conducted at the Hanover Park Treatment Plant in 1968. This work has been reported in detail by Lynam et al. (1)

Two methods of physical separation or removal of suspended solids were investigated, namely sand filtration using the Hardinge design rapid sand filter and microstraining using the Glenfield and Kennedy microstrainer. The Hardinge sand filter utilizes a silica sand with an effective size of 0.51 mm and a uniformity coefficient of 1.62. These filters are continuously cleaned by a traveling backwash mechanism so that no major down-time is experienced during backwash procedures. This unit permits about 90% of the filter bed to be in continuous operation, the remaining 10% being that portion being automatically backwashed by the travelling backwash mechanism.

Microstraining is a method of filtration in which a stainless steel fabric is used as a filtering medium. The microstrainer used at Hanover Park has a drum 10 feet in diameter and 10 feet long in which the fabric is mounted on the periphery of the revolving drum. The untreated water flows into the drum and radially outward through the microfabric, leaving behind the suspended solids removed by the fabric. The solids on the inside are carried upward, where a row of backwash jets flush them into a

(1) Tertiary Treatment at Metro Chicago by Means of Rapid Sand Filtration and Microstrainers

hopper mounted on the hollow axle of the drum. The drum speed and backwash pressure are automatically controlled by headloss across the drum. The maximum headloss is 6 inches; drum speed can vary from a minimum of 0.7 to a maximum of 4.3 rpm and correspondingly, the backwash pressures vary from 20 to 55 psi.

In the process of analyzing the large amounts of data collected it was discovered that plots of many parameters yield poor correlations and were not useful in evaluating the efficiency of the tertiary treatment process. However, a plot of suspended solids removal versus suspended solids loading in terms of lbs/ft^2/day yield a high degree of correlation. Consequently, this method of data presentation will be used to summarize the Hanover Park Operations.

RESULTS - SAND FILTRATION

The performance of the sand filter is shown in Figure 1. The regression equation is $y = 0.756X - 0.034$ with a correlation coefficient of 0.922. A family of curves may be constructed from the regression equation of Figure 1 to show a relationship between tertiary effluent quality in terms of suspended solids and hydraulic loadings. This family of curves (Figure 2), demonstrates the relationship between hydraulic loading and solids loading. The dashed line shows both an upper limit of 6 gpm/ft^2 for hydraulic loading and a solids loading limit of 0.65 lbs/ft^2/day to produce a tertiary suspended solids quality dependent upon the combination of secondary effluent quality and hydraulic loading. The same analysis was applied in terms of effluent BOD and the family of

FIGURE 1
CONTINUOUS FLOW SAND FILTER

SUSPENDED SOLIDS REMOVAL AT HIGH HEAD WITHOUT COAGULATION

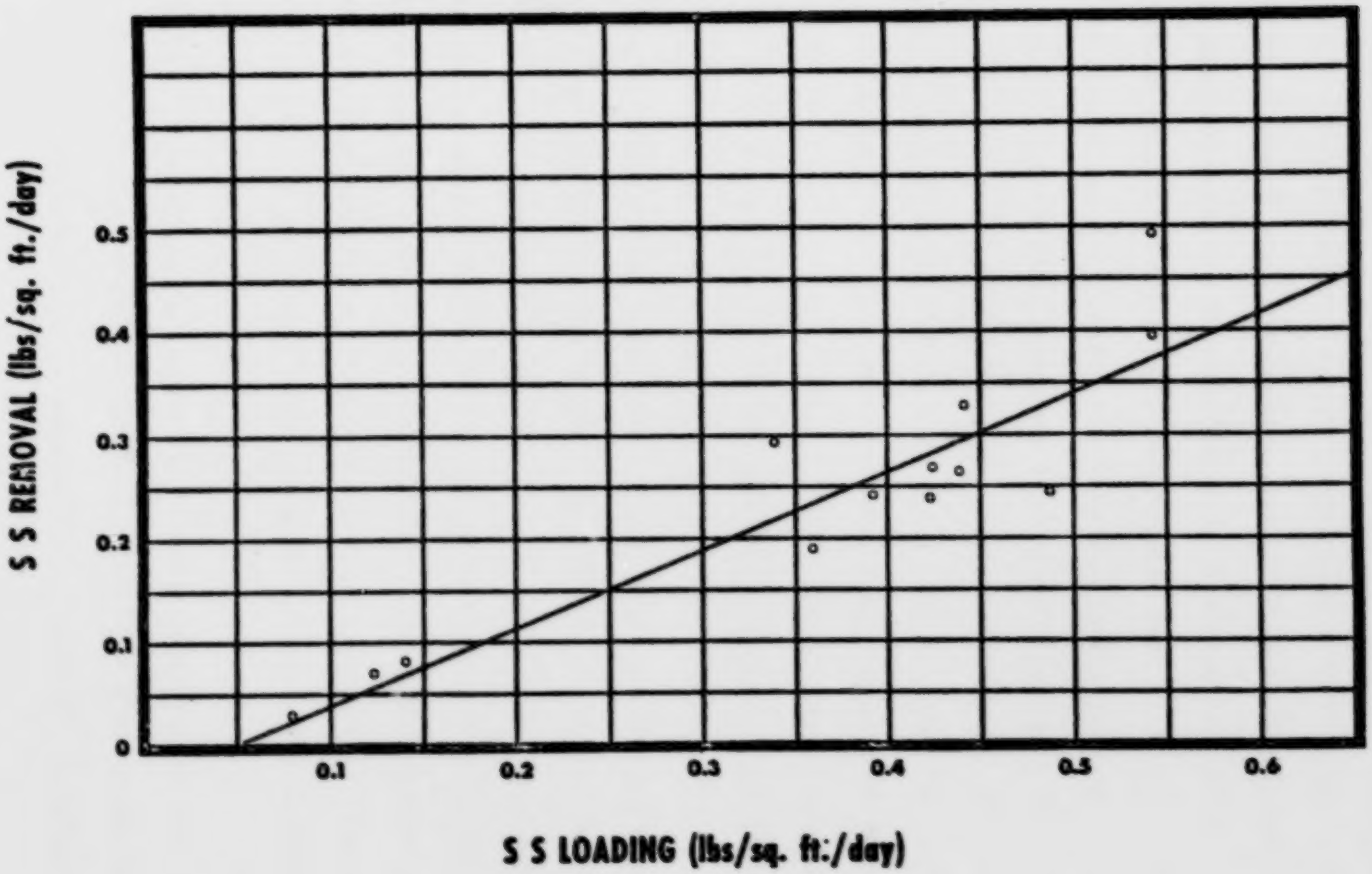


FIGURE 2
CONTINUOUS FLOW SAND FILTER

TERTIARY EFFLUENT SUSPENDED SOLIDS AS A FUNCTION OF HYDRAULIC LOADING AND SECONDARY EFFLUENT SUSPENDED SOLIDS HIGH HEAD BASED ON REGRESSION OF Y = 0.756X - 0.034

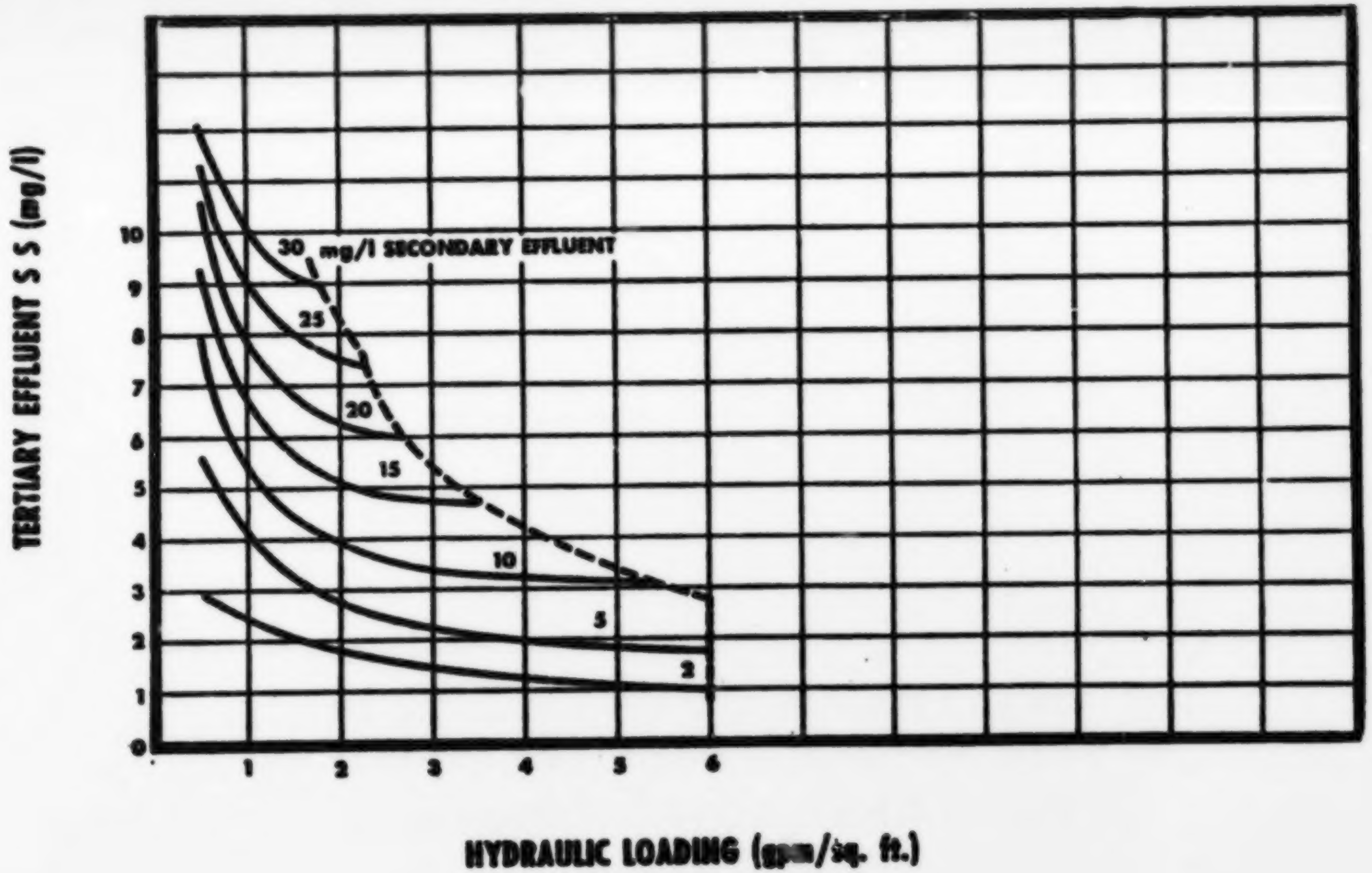
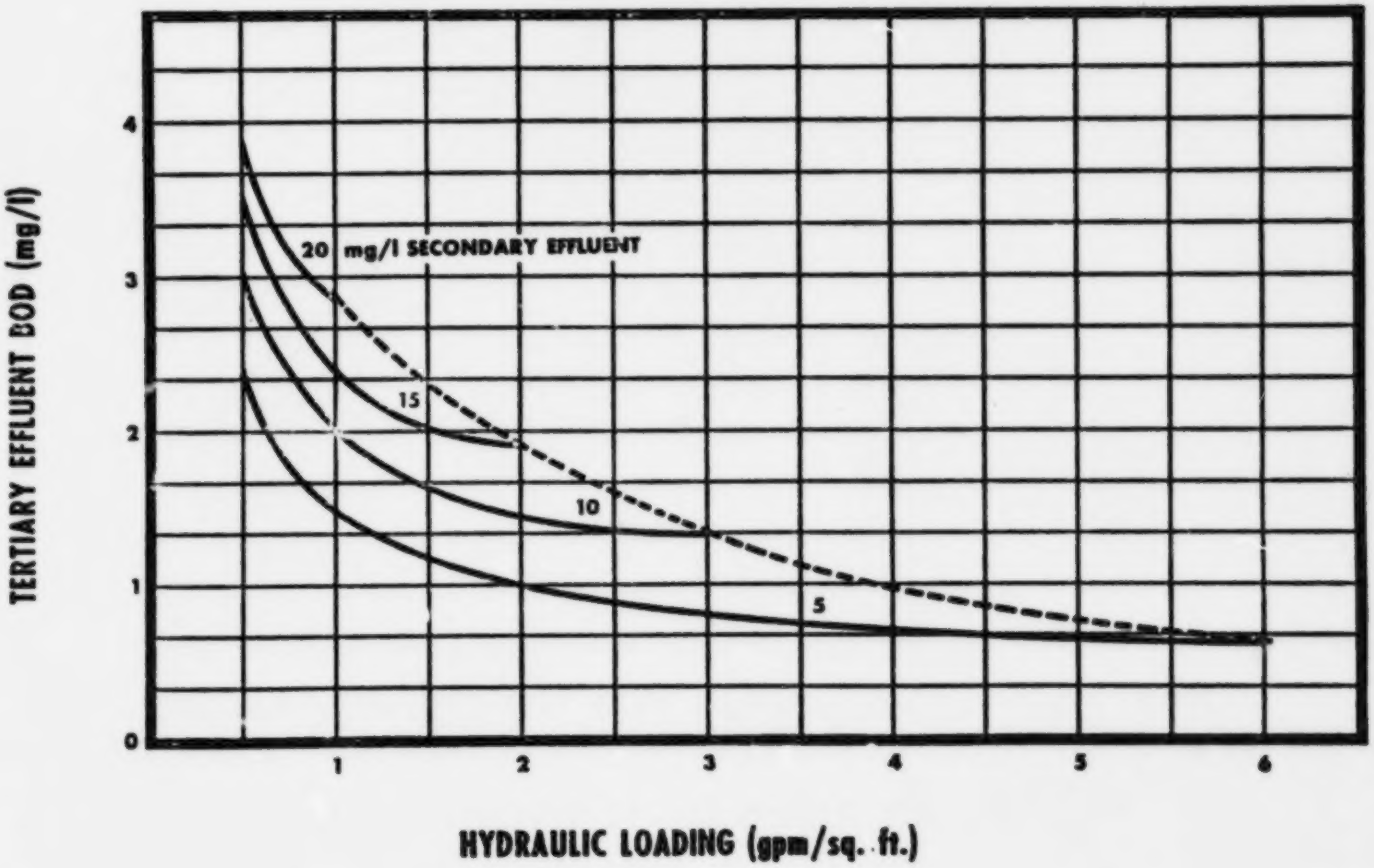


FIGURE 3
CONTINUOUS FLOW SAND FILTER

TERTIARY EFFLUENT BOD AS A FUNCTION OF HYDRAULIC LOADING AND SECONDARY EFFLUENT BOD BASED ON REGRESSION OF Y = 0.906X - 0.012



curves developed is shown in Figure 3.

MICROSTRAINER

Microstrainer performance at 10 percent idle speed (0.7-rpm) is given in Figure 4. The regression equation is $y = 0.739X - 0.011$ with a correlation coefficient of 0.922. The suspended solids removal at 0.7-rpm was significantly greater than at a 20 percent idle speed. It is believed that the slower speed allowed a greater build-up of solids at the surface of the microstrainer fabric thus yielding improved filtration. A family of curves, may be constructed from Figure 4 to show the relationship between tertiary effluent suspended solids and hydraulic loading. This family of curves is shown in Figure 5. The dashed line shows an upper limit of 6.5 gpm/ft^2 for the hydraulic loading and 0.88 lbs/ft^2/day for the pounds loading. A similar analysis was applied in terms of BOD and a family of curves developed as presented in Figure 6.

When one evaluates the overall efficiency of a physical separation process consideration must be given to the amounts of treated water used to backwash the unit after the maximum head is reached. Table 1 presents backwash consumption for the sand filter at flows of 1.0 through 6.0 gal/min/ft^2. The backwash exceeded 0.5% only once and that was at the maximum hydraulic loading. The percent of treated water used in backwashing the microstrainer increased to approximately 3 percent at 3.8 gal/min/ft^2 hydraulic loading.

FIGURE 4
MICROSTRAINER

SUSPENDED SOLIDS REMOVAL WITH IDLE SPEED AT 10% OF MAXIMUM

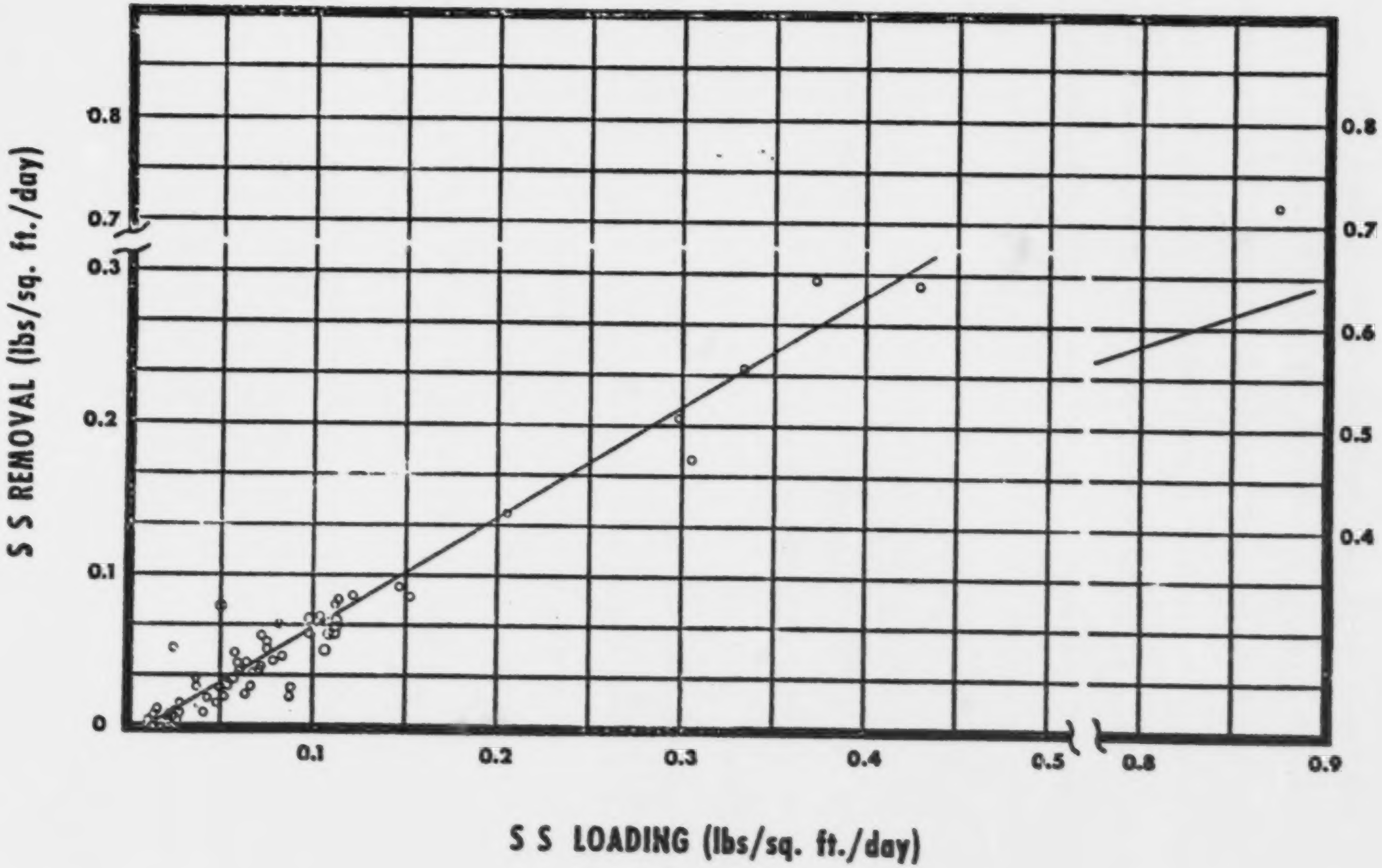


FIGURE 5 MICROSTRAINER
TERTIARY EFFLUENT SUSPENDED SOLIDS AS A FUNCTION
OF HYDRAULIC LOADING AND SECONDARY EFFLUENT SUSPENDED SOLIDS
BASED ON REGRESSION OF Y = 0.739X - 0.011

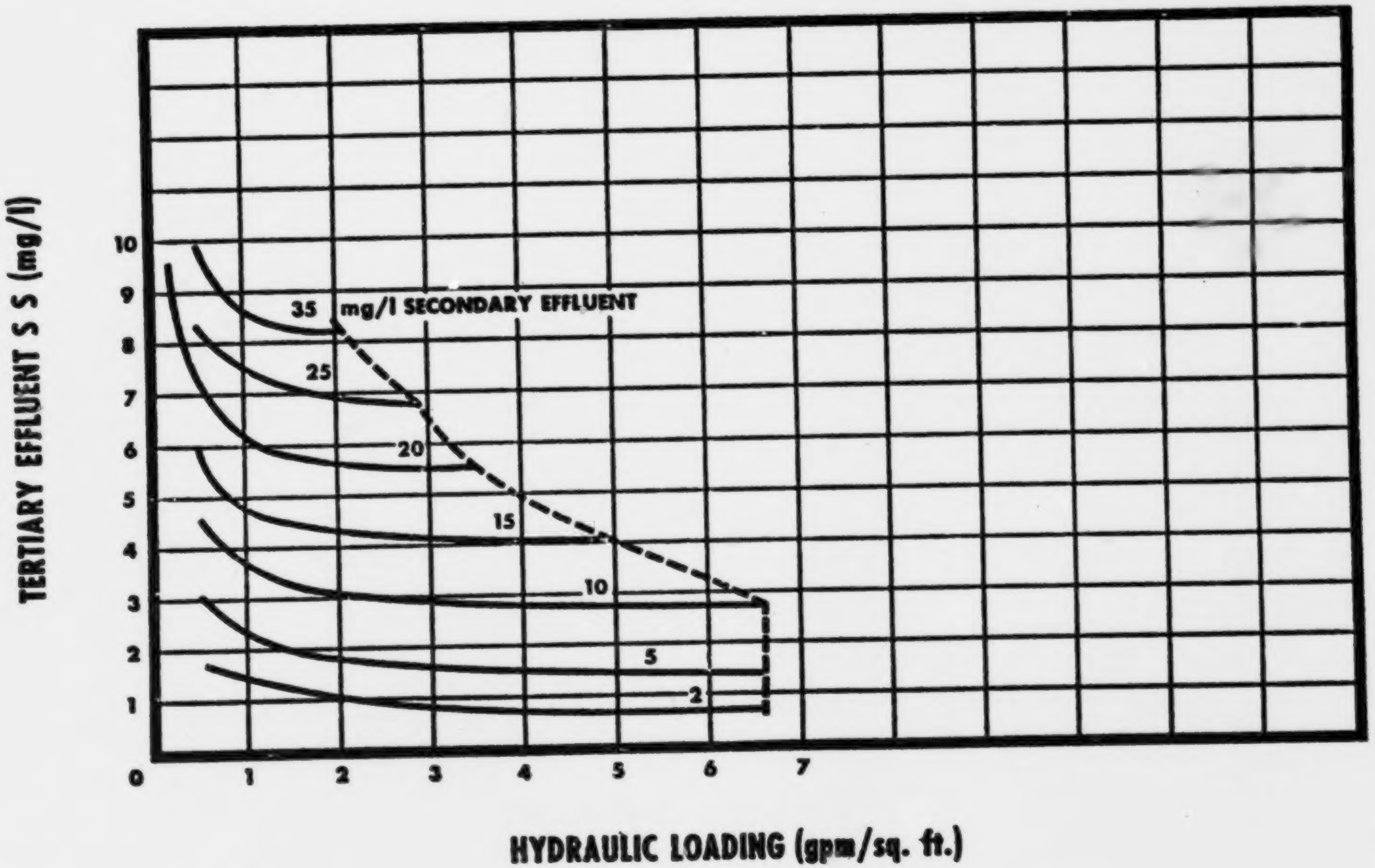
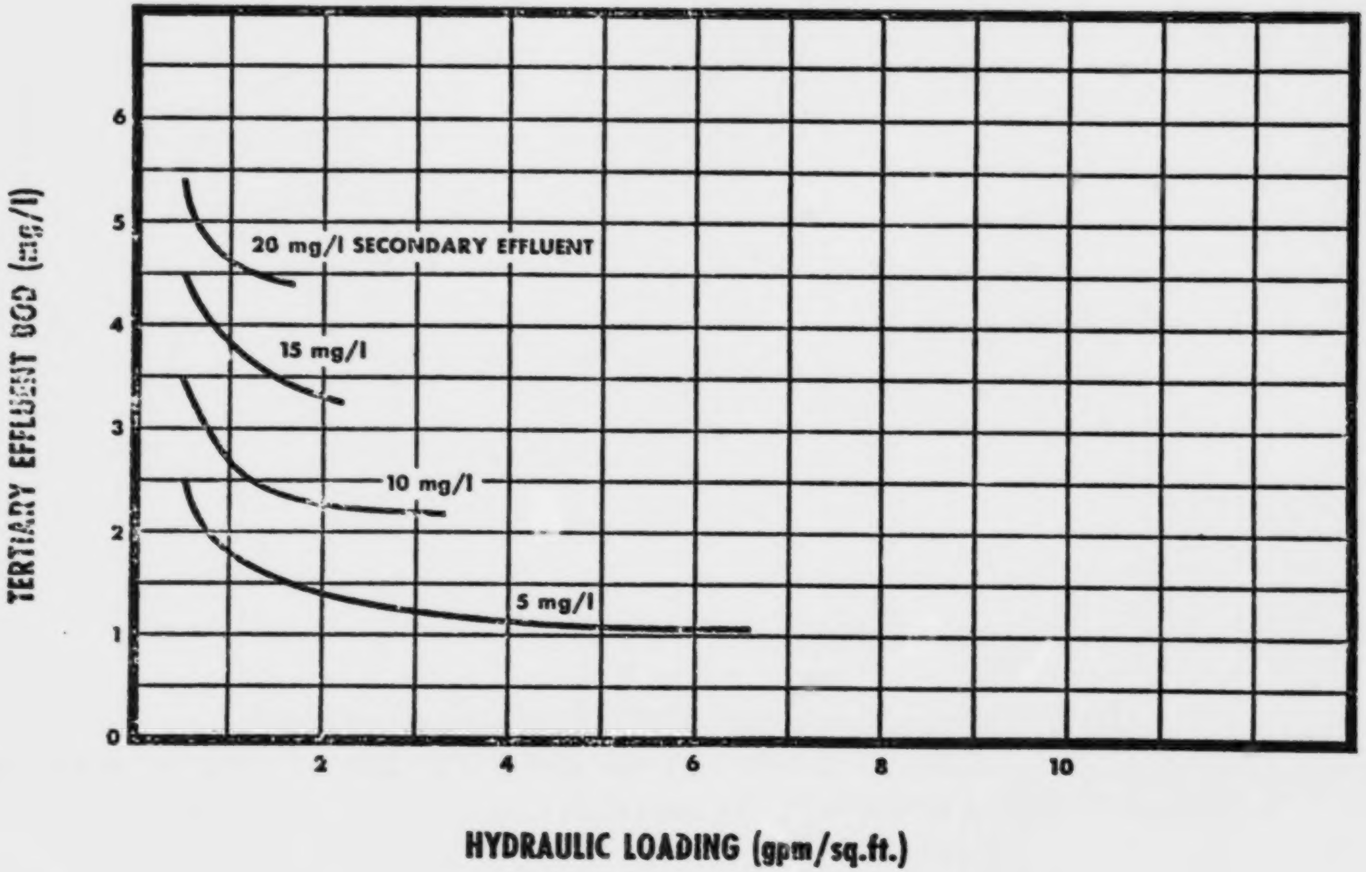


FIGURE 6
MICROSTRAINER

TERTIARY EFFLUENT BOD AS A FUNCTION OF HYDRAULIC LOADING AND SECONDARY EFFLUENT BOD. BASED ON THE REGRESSION OF Y = 0.736X + 0.006



METROPOLITAN SANITARY DISTRICT OF GREATER CHICAGO

TABLE 1

HANOVER PARK SAND FILTERS
% BACKWASH

FLOW Gal/(Min)(Sq Ft)	% BACKWASH
1	0.31
1.5	0.33
2	0.33
2.5	0.35
3	0.34
3.5	0.36
4	0.37
4.5	0.40
5	0.42
5.5	0.48
6	0.65

CONCLUSIONS

1. The maximum suspended solids loading to the sand filter was 0.65 lb/ft^2/day.

2. The maximum hydraulic loading to the sand filter was 6 gpm/ft^2.

3. With an influent of 18 mg/l of BOD and suspended solids, an effluent quality of 2.5 and 5.5 mg/l respectively can be obtained by the Hardinge sand filter.

4. The maximum suspended solids loading to the microstrainer was 0.88 lbs/ft^2/day.

5. The maximum hydraulic loading to the microstrainer was 6.5 gpm/ft^2.

6. With an influent of 18 mg/l of BOD and suspended solids an effluent quality of 4.0 and 5.0 mg/l respectively can be obtained for the Glenfield-Kennedy unit.

7. The sand filter backwash consumption was generally less than 0.5% up to the maximum flow, while the microstrainer reached 3% at the maximum flow.

PILOT PLANT STUDIES AT THE HANOVER PARK TREATMENT WORKS - EVALUATION OF 3 SAND FILTRATION DEVICES

INTRODUCTION

Following the studies conducted at the Hanover Treatment Works involving the use of the Hardinge continuous flow sand filter, it was the decision of the staff of the District that other sand devices should be evaluated. This was necessary because of the large amount of funds which would be required to meet the 1977 Illinois S.S. and BOD criteria (30 day average not to exceed 12 mg/l of S.S. and 10 mg/l BOD) and the necessity to select a treatment method which would yield the most cost effective solution to achieving such criteria.

The MSDGC decided 3 pilot sand filtration devices would be run in parallel and that this would enable direct comparison of these processes without accounting for differences in feed source.

An experimental area was set aside in the existing Hanover Tertiary treatment plant and three types of deep bed, high rate sand filtration pilot plants were set up. The three devices were a DeLaval upflow filter, a Neptune Microfloc mixed media filter and a Graver dual media pressure filter.

The objective of the program was to investigate the performance of each unit.

All of the units received a common influent, that is, secondary effluent from the Hanover Treatment Works. Direct simultaneous comparisons of effluent quality and operational parameters were therefore readily possible.

It should be noted here that all data presented is based upon filter runs, that is, the sampling was so arranged that influent and effluent samples were gathered between backwash cycles. Therefore influent and effluent analysis as well as filter loadings are based upon filter runs which were usually not of a 24-hour duration. Loading and removal figures are given in lbs/ft^2/filter run and influent and effluent samples are not generally for 24-hour periods.

Utilizing influent and effluent analysis as well as the loading and removal values based upon 24-hour samples is justified for the previously discussed full scale Hanover studies, because both the Hardinge sand filter and the Glenfield-Kennedy Microstrainer backwash virtually continuously. Attempting to orient analysis of data upon runs for the microstrainer and Hardinge filter is not theoretically possible while the Hardinge filter backwashes many times during a given 24-hour period.

Orienting analysis of data upon filter run for the 3 pilot sand filters is justified due to the batch nature of the 3 devices undergoing study. Loading and removal values based upon filter runs yield the total lbs of S.S. per square foot applied or removed between backwashes and are more meaningful and correlatable with other operational and performance criteria. Samples gathered during a filter run while often not actually occurring during a calendar day should be comparable with samples composited on a 24-hour calendar day.

EXPERIMENTAL APPARATUS

1. The DeLaval Filter

The filter used was manufactured by the DeLaval Separator Company and was an "Upflo Immedium filter Model OT-3". The filter vessel was approximately 3 feet in diameter and 13 feet deep. The filtering media was silica sand and gravel, with the filter bed being approximately 7 feet deep. The bed had an effective surface area of 6.75 square feet and at a maximum flow rate of approximately 6 gpm/ft^2, the unit can filter about 58,000 gallons daily.

As can been seen in Figure 7, the flow through the filter was upward, with the influent entering the unit through a large number of nozzles located in a distribution plate at the bottom of the filter vessel. The filter media consisted of four layers, with two of the layers being gravel and two of the layers being sand. The two gravel layers served as support for the sand layers, as well as distributing the flow uniformly and thereby reducing short circuiting. The first gravel layer which was at the bottom, was a 4 inch layer consisting of coarse gravel (1-1/4 to 1-1/2 inches). The second layer of gravel was 10 inches thick and contained gravel (3/8 to 5/8 inches). On top of the gravel layers was a 12 inch layer of coarse sand (2 to 3 mm). The filter bed was held in place by a grid, which was buried near the top of the 60 inch fine sand layer. This filter was designed to utilized the entire depth of the filter for filtration and solids storage.

The backwash cycle was accomplished as follows: initially the filtration cycle was terminated when the head loss through the filter reached a preset level, which was usually 14 psi. With the filtration cycle terminated, the backwash cycle was then initiated

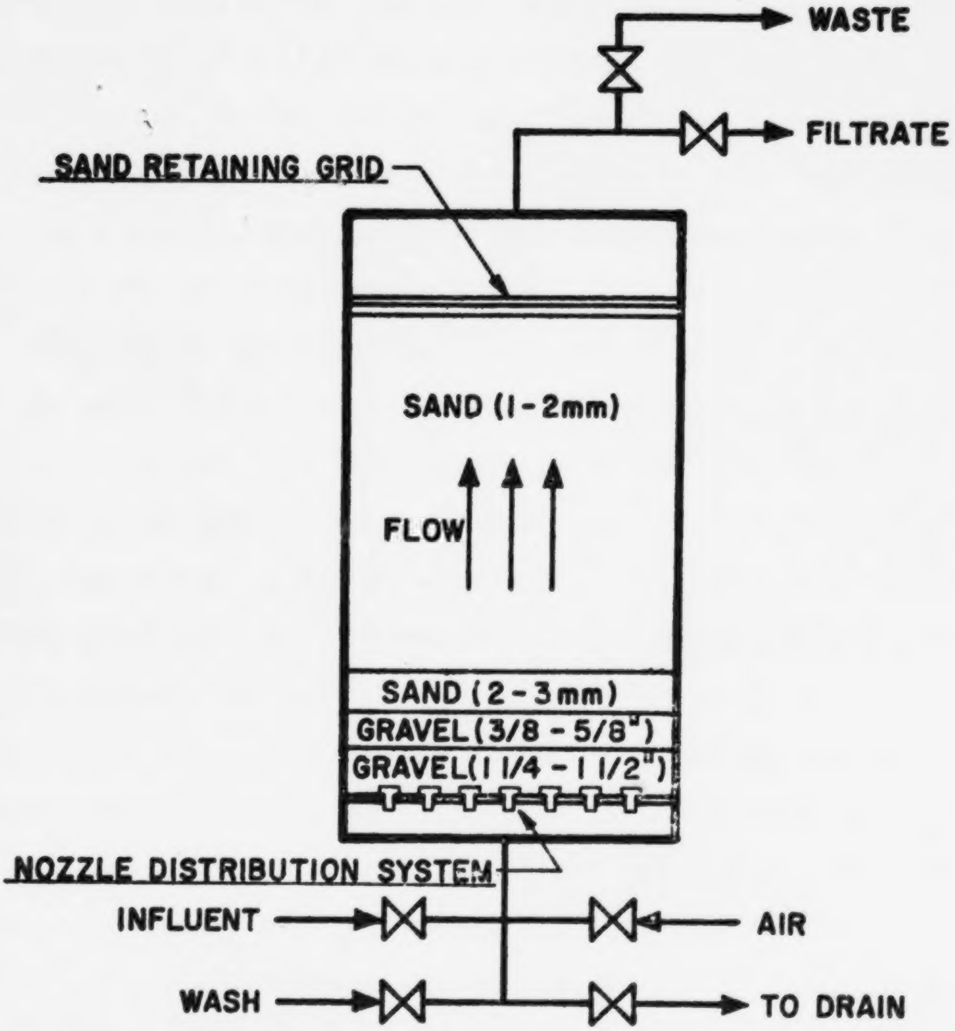


FIGURE 7

DIAGRAM OF THE DeLAVAL FILTER

by first draining the filter bed. The filter was then fluidized by forcing air through the filter at low pressure (5 to 10 psi). After 3 minutes of air flow, the filter was flushed with water at a rate of about 10 - 13 gpm/ft^2. After approximately ten minutes of flushing the bed was allowed to settle for five minutes, with the largest particles settling to the bottom and the finer particles to the top.

2. The Neptune Microfloc Unit

The unit used was manufactured by Neptune Microfloc Inc. and was a "Reclamate SWB-27A." The principal tank was 5 feet square and 6 feet deep, and was divided into three compartments: a flocculation chamber, a settling chamber, and a filter chamber. The settling chamber contained settling tubes and the filtering media consisted of anthracite coal, silica sand, garnet, and gravel. The filter bed was 5 feet deep and had a surface area of 4 square feet, and a maximum flow through rate of 10 gpm/ft^2 or about 58,000 gpd. The backwash storage tank was approximately 5 feet in diameter and 7 feet deep. Flow through the filter was regulated by an effluent pump and an effluent rate control valve which was operated by a level transmitter positioned above the bed.

The filtering media consisted of from top to bottom: a 30 inch layer of anthracite coal (1.2 to 1.3 mm), a 12 inch layer of silica sand (0.8 to 0.9 mm), a 6 inch layer of garnet (0.4 to 0.8 mm), and a 3 inch layer of support gravel 1.5 to 2.0 inches, and at the bottom the entire bed rests on a 12 inch layer of gravel (1/2 to 2 inches).

As can be seen in Figure 8 the "Reclamate SWB-27A" could be operated in several different modes. The complete flow pattern included chemical addition followed by flocculation, settling, and removal. However as shown in Figure 8 if chemical addition was not desired, the flocculation and settling chambers could be bypassed. Therefore, the appropriate mode of operation can be selected on the basis of the quality of the influent wastewater, the nature of the suspended solids, and degree of tertiary treatment desired within the performance limits of the unit. The studies noted in this paper will only include those tests when the flow bypassed the flocculation and settling chambers and no coagulants were utilized.

When the head loss through the unit increased to the setting on the vacuum switch located between the filter and the effluent pump, the backwash cycle was initiated. As shown in Figure 8, both the filter and the settling chamber were cleaned during the backwash cycle. The backwash flow was 20 gpm/ft^2 and the volume of water required for a backwash was approximately 650 gallons. Settling after backwash restored the anthracite coal, silica sand, garnet and gravel layer to their proper positions in accordance with their density differences.

3. The Graver Filter

The filter manufactured by the Graver Water Conditioning Company was a "Monoscour Filter," and the Wastewater was pumped down and through the filter. A diagram of the Graver filter is shown, in Figure 9. The filter vessel was 22 inches in diameter and 7 feet 6 inches in height. The filtering media consisted of

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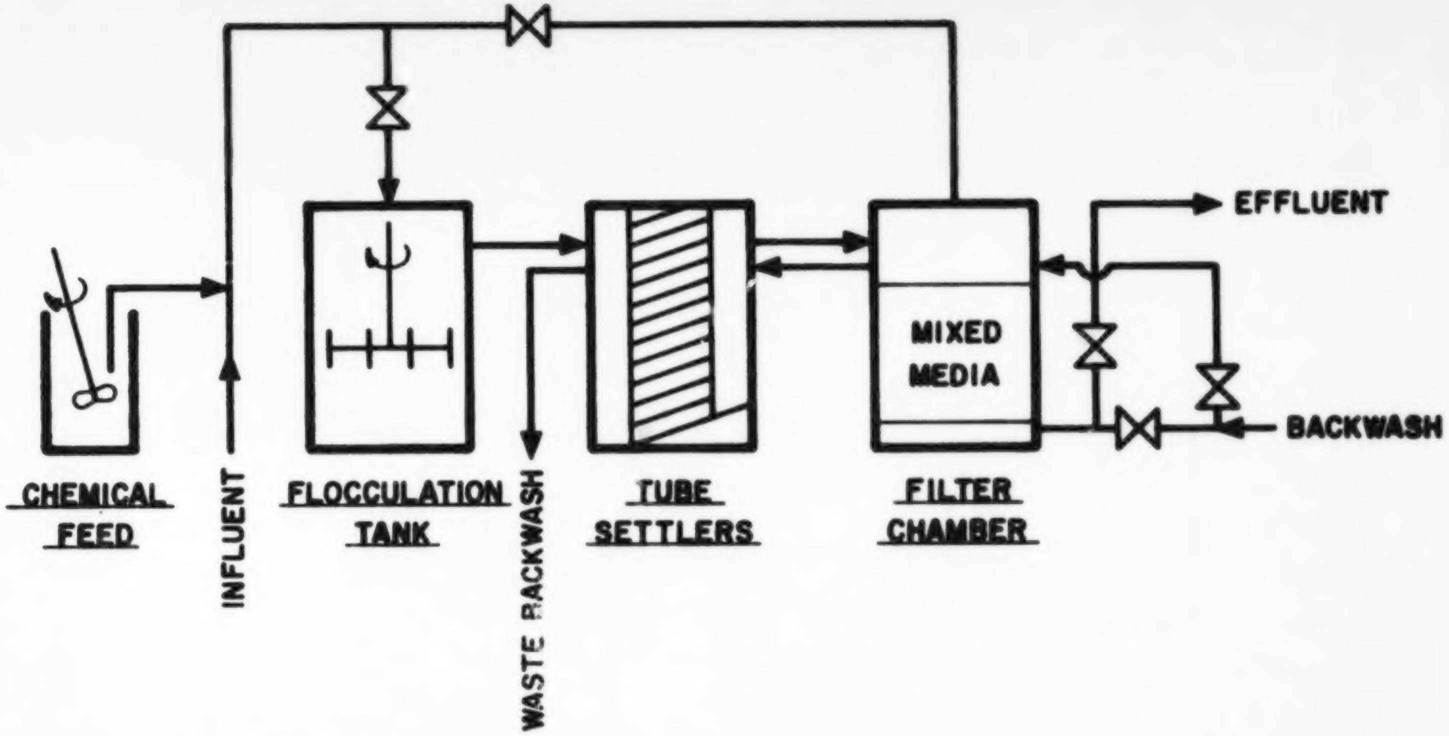

FIGURE 8

FLOW DIAGRAM OF THE NEPTUNE MICROFLOC UNIT

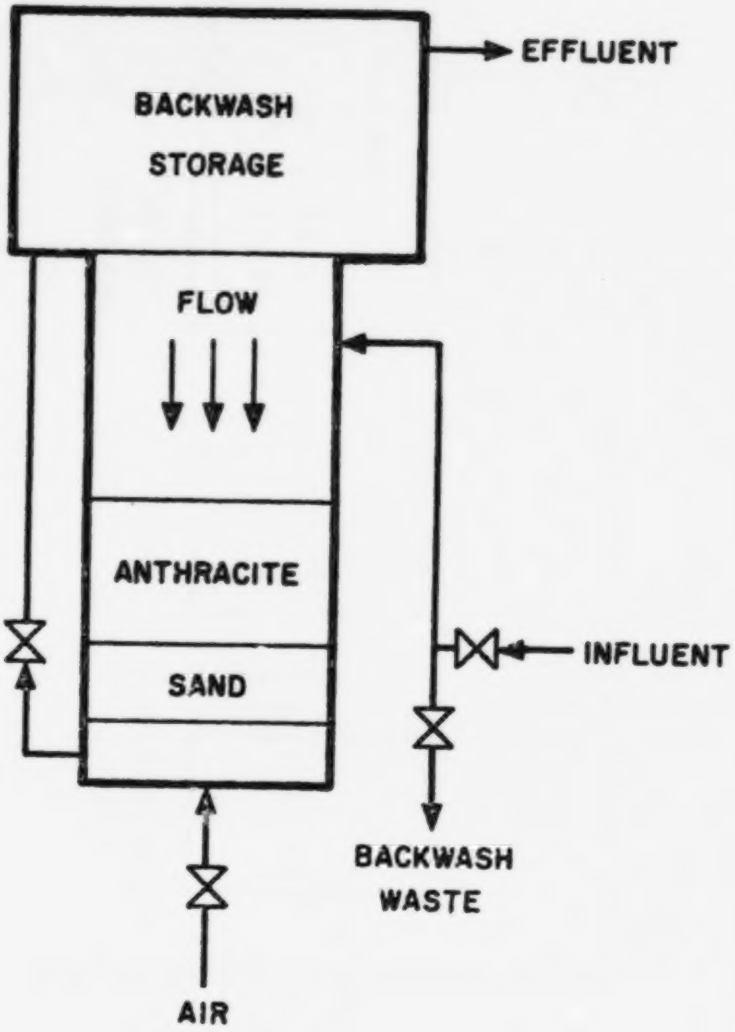

FIGURE 9

DIAGRAM OF THE GRAVER FILTER

anthracite coal and silica sand, with the filter bed being approximately 3 feet deep. The bed had an effective surface area of 2.65 square feet and at a maximum flow rate of 11.5 gpm/ft^2 the unit was capable of filtering about 43,000 gallons daily. As shown in Figure 9, the backwash storage compartment (6 feet in diameter and 5 feet in height) was positioned directly above the filter vessel.

A combination of anthracite coal and silica sand were used during the test period with the size of the anthracite being 1.0 to 1.4 mm and the size of silica sand being 0.6 to 0.7 mm.

As in the case of the other filters, the filtration cycle was automatically stopped when the influent pressure to the filter reached a preset level. The backwash cycle begins with the bed being initially drained. After the bed was drained, the filter bed was then air scoured for 5 minutes at a rate of 15 scfm at 5 psi. Following the air scouring the filter bed was allowed to settle for 6 minutes, after which the filter bed was backwashed for 5 minutes at a rate of about 15 gpm/ft^2. The total volume of water used during backwash was about 200 gallons. After backwashing the filter bed, it was allowed to settle, with the sand settling below the anthracite coal because of its greater density.

RESULTS

DeLaval Filter

In Table 2 are contained results of the operation of the DeLaval filter at three different flow rates.

Generally speaking, as the loading increased so did backwash frequency and volume. Based upon the data collected, it appears that the unit is capable of treating loadings up to 1.24 lbs/ft^2/day with backwash rates of less than 4%.

It also appears, from the data collected, that the DeLaval filter is capable of handling hydraulic loadings of 4-5 gpm/ft^2 with the effluent solids being about 5 - 7 mg/l and the effluent BOD about 5 - 9 mg/l.

Neptune Microfloc

The results obtained for the Neptune Microfloc unit are represented in Table 3. The unit was tested at flow rates of 2 and 5 gpm/ft^2. An increase in hydraulic loading, as compared to the DeLaval unit, caused a reduction in loading. As one would expect, backwash frequency and, therefore, backwash usage increased with increased hydraulic loading. Effluent BOD and S.S. were between 4 to 6 mg/l and appeared to be independent of hydraulic loading. The Neptune Microfloc unit appears to be capable of achieving excellent effluent quality (less than 6 mg/l of S.S. and BOD) at hydraulic loadings of 4 gpm/ft^2 with backwash volumes less than 3%.

Graver Pressure Filter

Figure 10 presents data on loading and removals of the Graver

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TABLE 2

OPERATION OF THE DE-LAVAL FILTER AT VARIOUS HYDRAULIC LOADINGS

HYDRAULIC LOADING, GPM/SQ FT	4	4.7	5
Influent S.S., mg/l	12	14	14
Effluent S.S., mg/l	7	7	5
S.S. Removal, %	44	54	62
Influent BOD, mg/l	7	17	18
Effluent BOD, mg/l	5	9	5
BOD Removal, %	26	48	72
Length of Filter Runs, Hrs	43.0	37.0	35.4
S.S. Loading, lb/sq ft/filter run	1.04	1.23	1.24
S.S. Removal, lb/sq ft/filter run	0.45	0.67	0.70
Backwash Usage, %	2.5	2.7	3.7

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TABLE 3

OPERATION OF THE NEPTUNE MICROFLOC UNIT TREATMENT OF SECONDARY EFFLUENT

HYDRAULIC LOADING , GPM/SQ FT	2	4
Influent S.S., mg/l	14	16
Effluent S.S., mg/l	4	4
S.S. Removal, %	73	73
Influent BOD, mg/l	23	25
Effluent BOD, mg/l	6	4
BOD Removal, %	74	85
Length of Filter Run, hours	106.3	27.2
S.S. Loading, lb/sq ft/filter run	1.53	0.86
S.S. Removal, lb/sq ft/filter run	1.12	0.64
Backwash Usage, %	1.3	2.5

Pressure Filter at flow rates from 2.2 to 9.0 gpm/ft^2. In general, loading and removal were closely grouped over the entire range of flow rates. The highest loading was achieved at a flow rate of 9.0 gpm/ft^2 where 1.9 lbs/ft^2 was achieved at that flow rate.

Figure 11 presents results for backwash requirements of the Graver filter. Clearly, backwash usage remained fairly constant over the range of flow rates. This would seem to correlate with the loading values given above in that if the loadings per filter run remain consistent, backwash usage should also be so. In general, backwash usage was low usually being less than 1% for all flow rates tested.

Figure 12 presents the effluent quality for the Graver filter. In general, for all hydraulic loadings, effluent solids range from 5 to 10 mg/l while effluent BOD values range from 6 to 8 mg/l. Apparently, effluent quality remains independent of hydraulic and S.S. loading.

The Graver filter appears capable of handling hydraulic loads of up to 9 gpm/ft^2 and S.S. loadings up to 1.9 lbs/ft^2 and achieve effluent quality averaging about 6 mg/l of S.S. and BOD.

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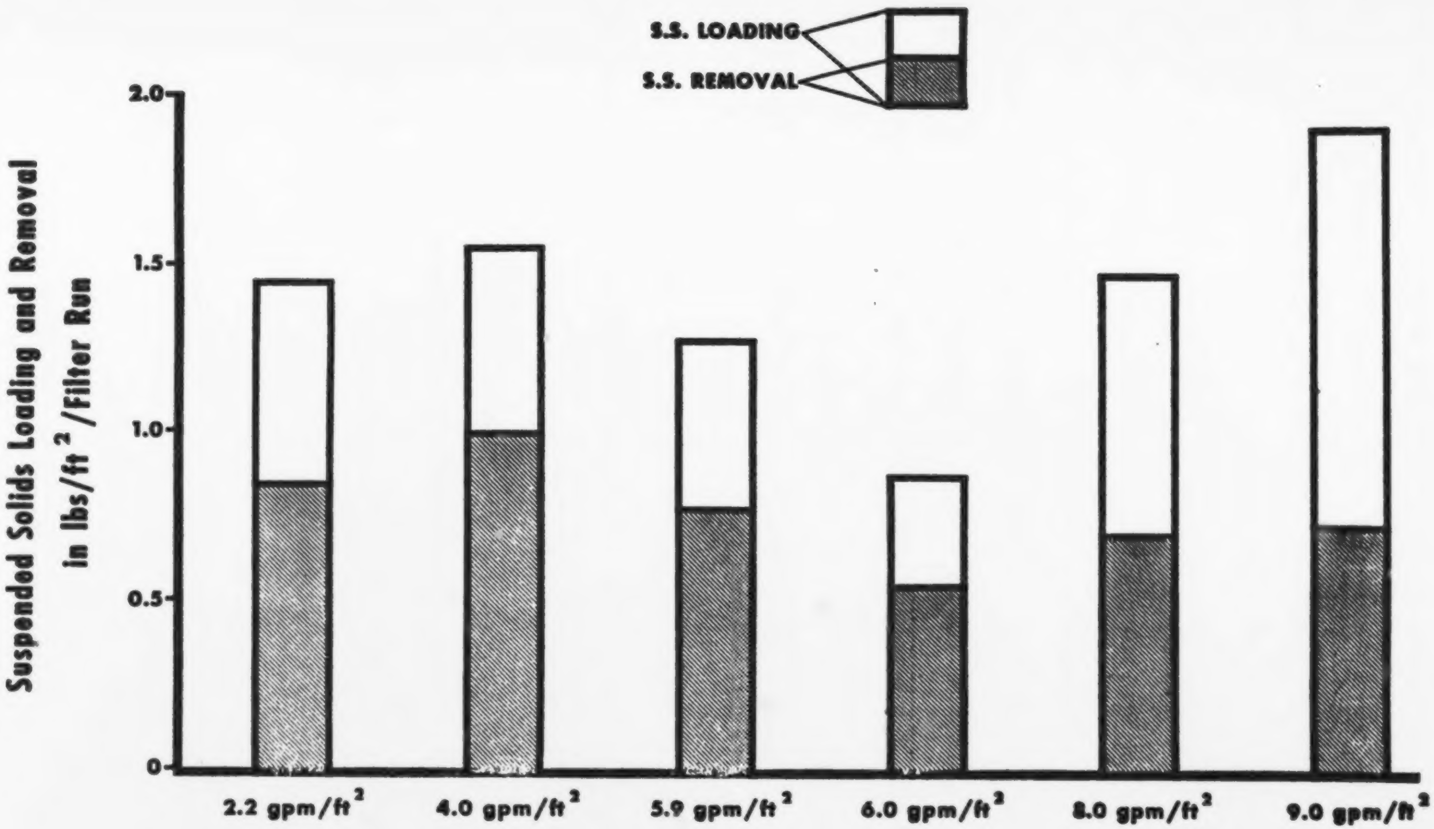
FIGURE 10

SUSPENDED SOLIDS LOADINGS AND REMOVALS FOR THE GRAVER FILTER



THE METROPOLITAN SANITARY DISTRICT OF GREATER CHICAGO

FIGURE 11

BACKWASH USAGE FOR THE GRAVER FILTER

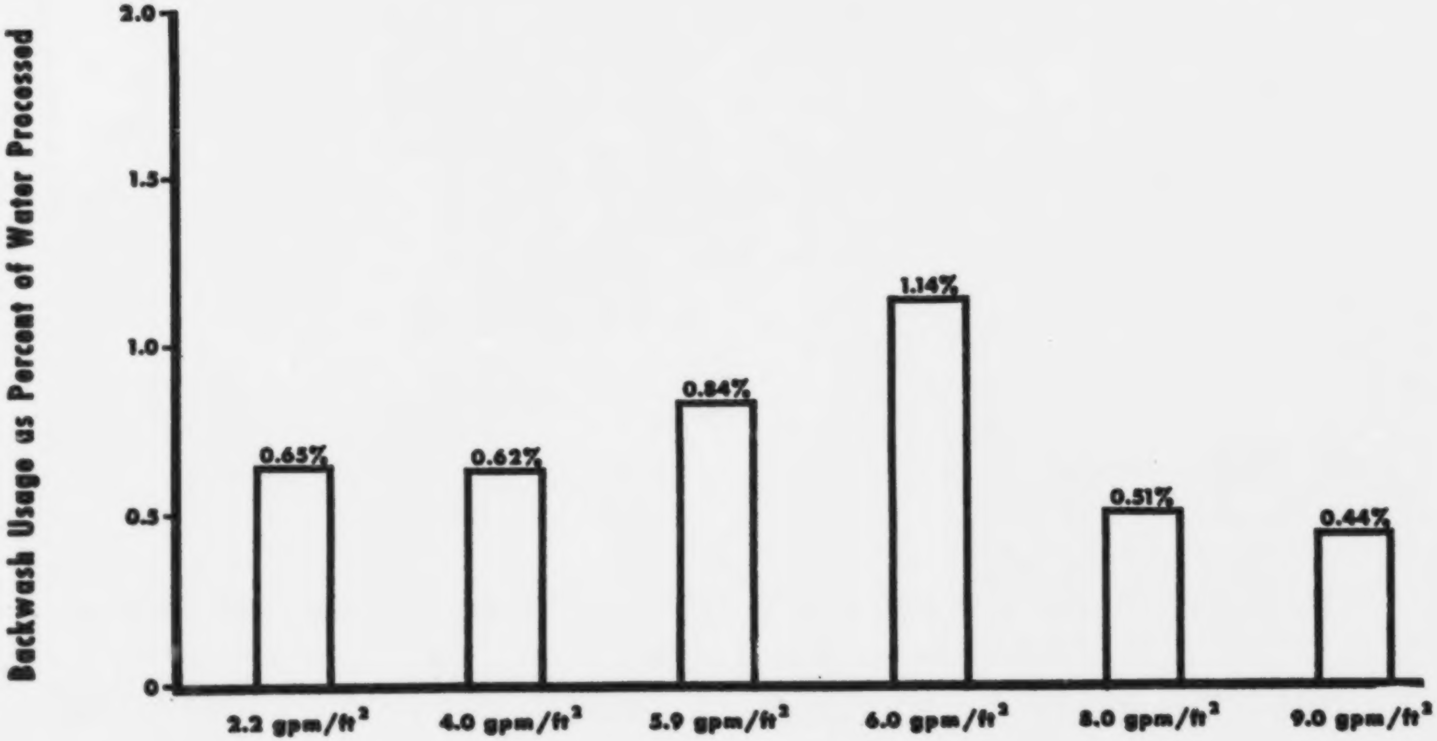


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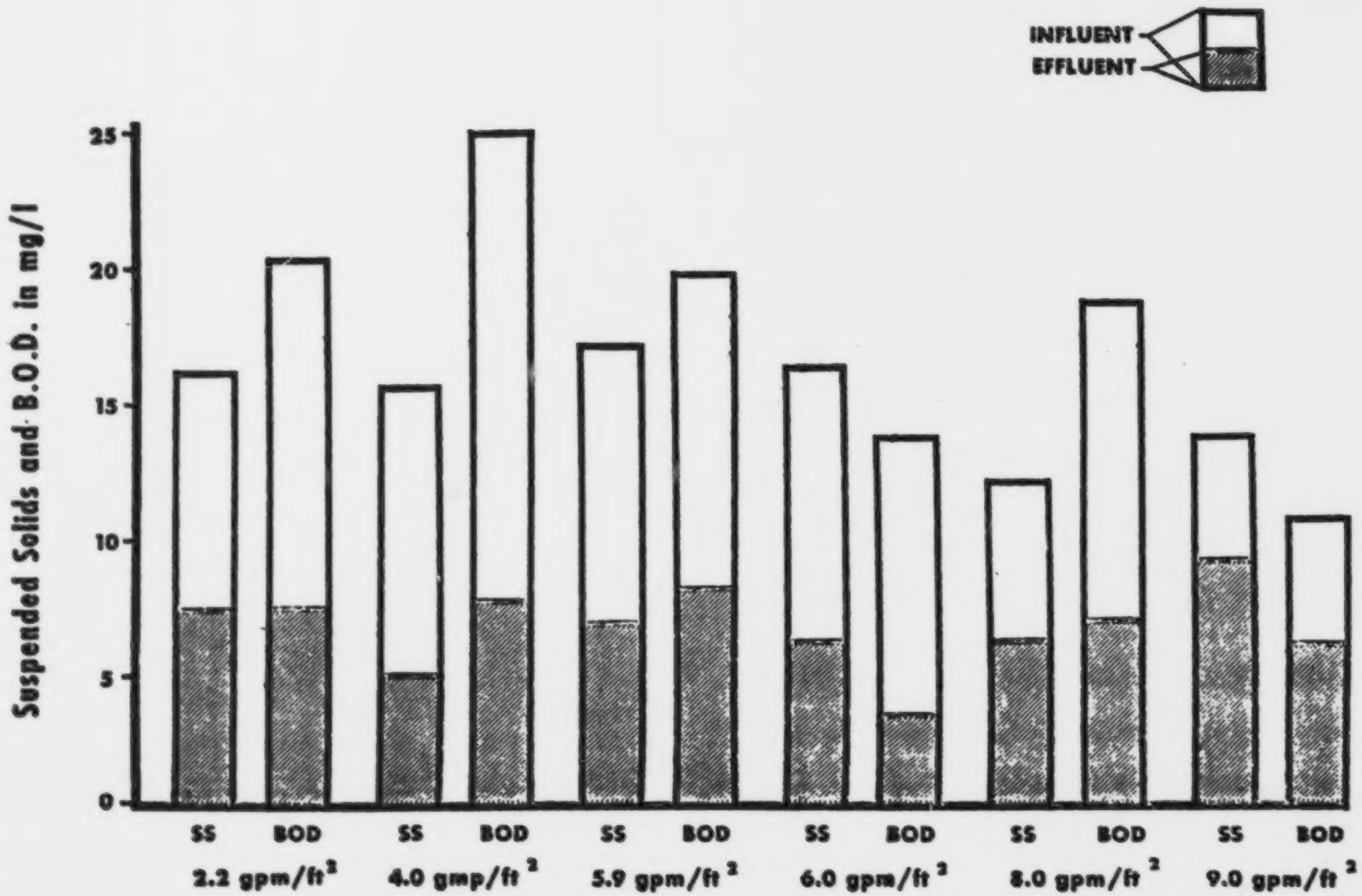
FIGURE 12

INFLUENT AND EFFLUENT SUSPENDED SOLIDS AND B.O.D. FOR THE GRAVER FILTER



CONCLUSIONS

From the results given on the previous pages, the following may be concluded from the pilot plant studies conducted in the Hanover Tertiary Building.

1. All the filtration units tested appeared to be capable of producing effluent S.S. and BOD of less than 10 mg/l.

2. The Graver pressure filter was capable of operating at higher hydraulic loadings (9.0 gpm/ft^2) and S.S. loadings (1.9 lbs/ft^2) than the other units tested.

3. The Neptune Microfloc unit was capable of achieving the best effluent quality (less than 6 mg/l of S.S. and BOD).

EVALUATION OF 15 M.G.D. RATED CAPACITY MICROSTRAINER AT THE NORTH SIDE TREATMENT WORKS OF THE METROPOLITAN SANITARY DISTRICT OF GREATER CHICAGO

INTRODUCTION

Following the sand filtration and microstrainer studies at the District's Hanover Works, the staff of the MSD concluded that the microstraining concept held promise for its North Side Treatment Works. This was based mainly upon the low land area requirements of the microstrainer and the limited expansion space available at the North Side Plant.

Although the microstrainer studied at the Hanover Works performed adequately, it was considered to be too small for practical consideration at the North Side Sewage Works. The microstrainer tested at Hanover (10' diameter drum, 10' long) was rated at 2.9 mgd based upon the studies conducted at Hanover. Clearly, for a 330 mgd (dry weather flow) plant, such a size unit would produce an installation with far too many individual units. Therefore, the MSDGC decided that single units whould have a least a 15 M.G.D. capacity for the North Side Facility.

Because no manufacturer had produced a microstraining device of 15 M.G.D. capacity, the MSD decided to issue a performance type contract for only one such unit for the North Side Plant. Commensurate with standard bidding procedures utilized by the District, a contract was issued to the low bidder for such a unit and a microstrainer constructed at the North Side Facility.

The facility supplied to the District was a drum 40 ft. long by 10 ft. in diameter. Unlike the microstrainer at Hanover, this unit receives influent from both ends of the drum. In addition, the filtering fabric was not placed flat on the outer surface of the drum, but in a corrugated pattern. The manufacturer stated that this was done to provide as much filtering area as possible for the drum size supplied.

The unit constructed at North Side was manufactured by the Crane-Cochrane Co. and utilized a Mark O stainless steel fabric with triangular openings of 23 X 45 X 45 microns (160,000 openings to the inch). Again, the fabric traps the solids and rotates with the drum to bring the fabric under wash water sprays which wash the solids into hoppers for gravity removal to disposal. After the fabric is washed, it passes under ultraviolet lights to inhibit bacteriological growths. With the corrugated arrangement of the fabric on the drum, it was possible for the manufacturer to place the fabric in removable sections.

The drum speed and backwash pressure can be automatically controlled by head loss through the drum. When a set head loss is exceeded, the drum speed and backwash pressure are increased until the set head loss is restored. The unit is built for maximum head losses of 6 inches, drum speeds of about 1 to 5 rpm and backwash pressure of about 35 to 37 psi.

The entire unit was housed in a heated all weather facility. Federal funding for performance tests on the unit were obtained and the unit began shakedown runs in November, 1971.

The unit was tested for flow capacity, that is its ability to filter secondary effluent from the North Side Plant at flow rates approaching 15 MGD. Operation of the unit consisted of achieving, as much as practicable, a head loss of 6 inches within the constraint of the unit's drum speed and backwash pressure.

RESULTS

In Figure 13 is presented a cummulative frequency of the loadings (lbs/ft^2/day) to the unit during the period of flow capacity testing. The unit depicts an ability to take solids loadings as high as 1.0 lb/ft^2/day. As noted previously for the Hanover studies, it is believed that this expression enables direct comparison between various types of units and sizes. Clearly, such an expression has value since filtration devices are designed primarily to remove S.S. and have an ability to remove only a certain total weight per unit area given a specific head loss and type of S.S.

In Figure 14 is presented a cummulative frquency curve of the flows (24 hour) achieved by the microstrainer for the loadings given in Figure 13. Clearly, the microstrainer rarely achieved the 15 MGD capacity envisioned for it and actually produced flows averaging (50 Percentile) 7.6 mgd.

Plotted in this same cummulative frequency chart are the flows expressed as gpm/ft^2. It can be seen that the unit achieves flows averaging 3.5 gpm/ft^2 for the solids loading depicted in Figure 13.

It should be noted here that the surface area of the fabric was taken to be total fabric area or the total area of the microstrainer farbir taking into account its corrugated placement on the drum.

In Figure 15 is presented cummulative frequency charts of the influent and effluent S.S. for the unit during the same flow

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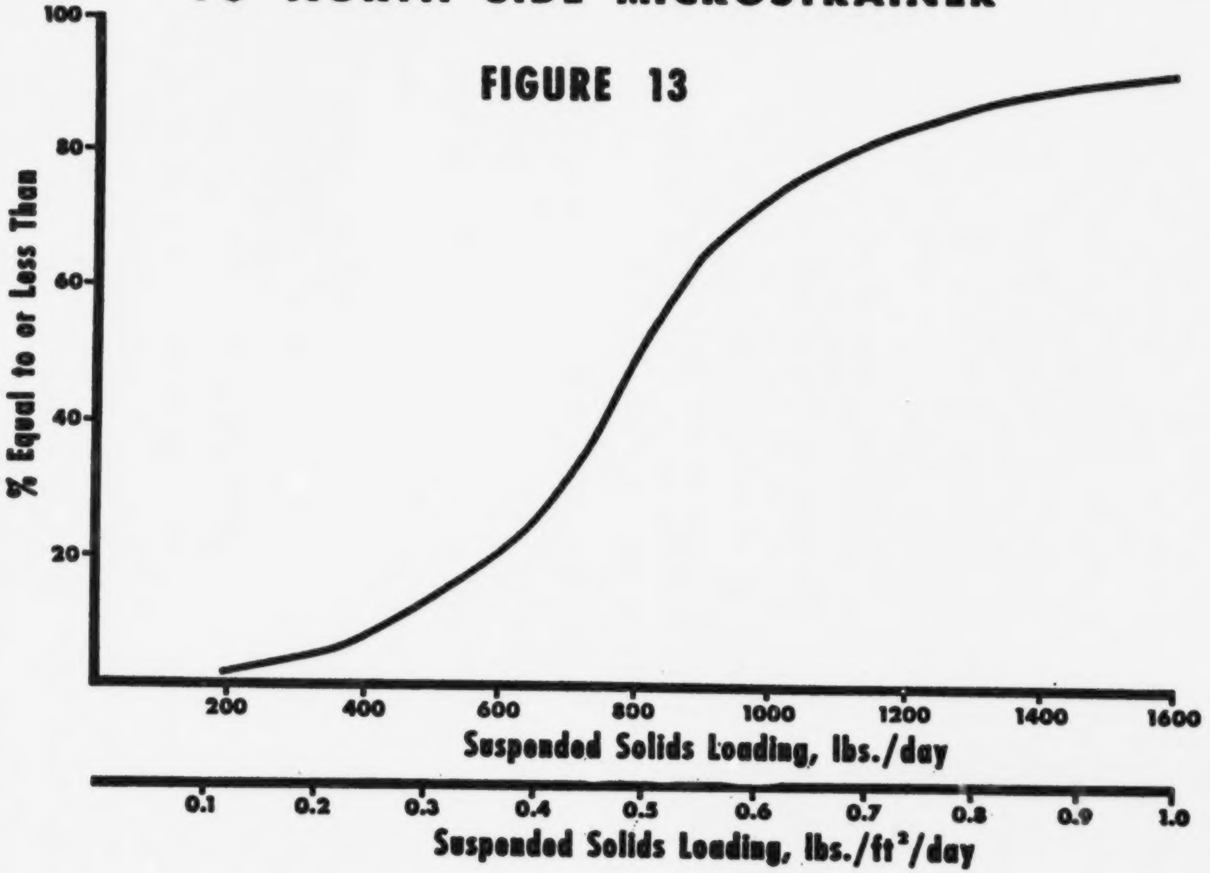
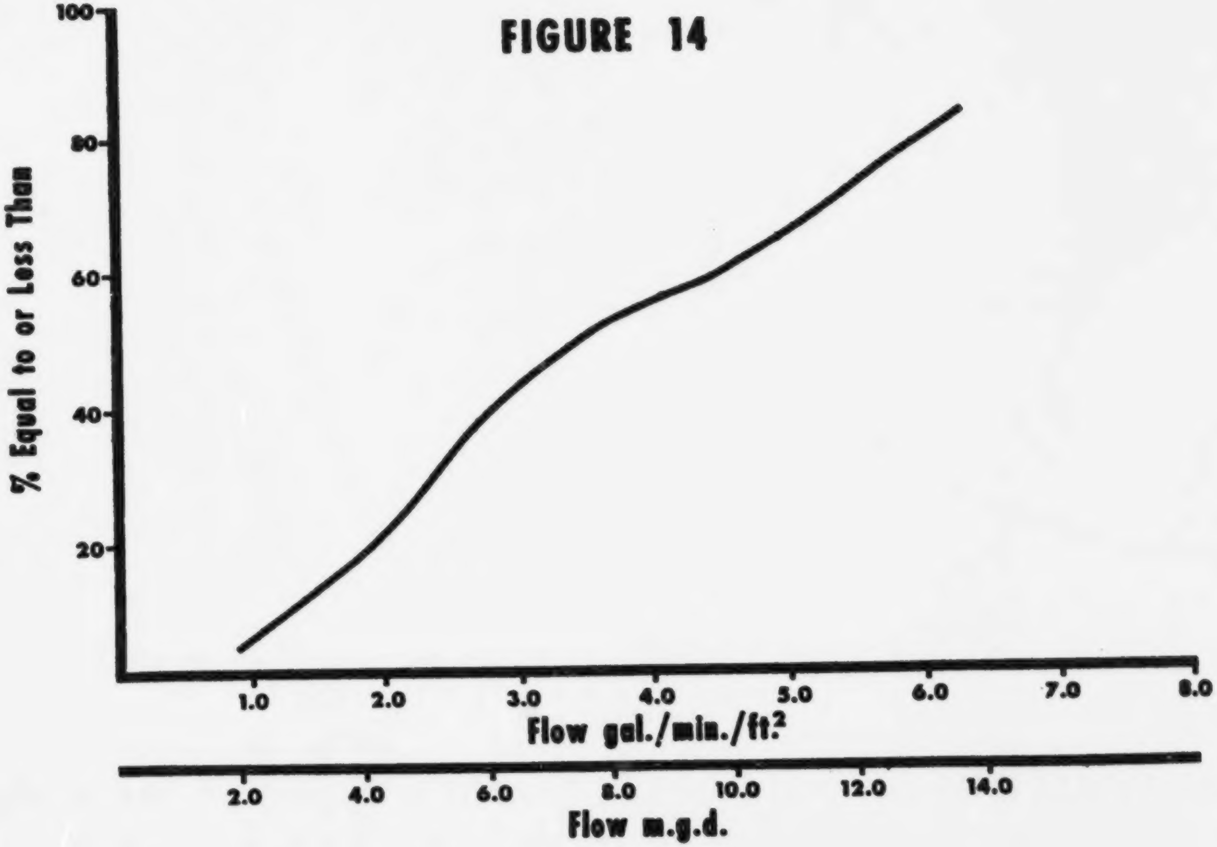
CUMULATIVE FREQUENCY DISTRIBUTION OF INFLUENT SUSPENDED SOLIDS LOADING TO NORTH SIDE MICROSTRAINER

FIGURE 13



THE METROPOLITAN SANITARY DISTRICT OF GREATER CHICAGO
CUMULATIVE FREQUENCY DISTRIBUTION OF FLOW THROUGH NORTH SIDE MICROSTRAINER

FIGURE 14



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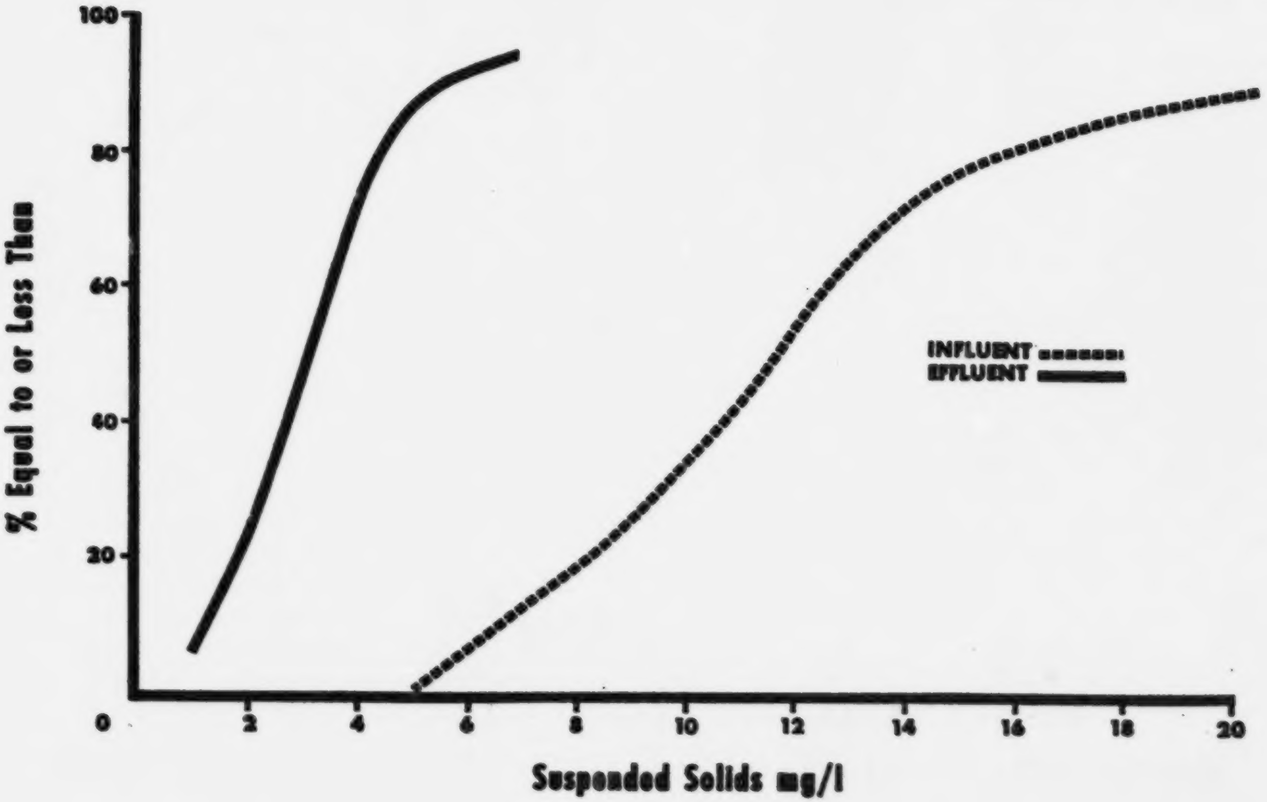
CUMULATIVE FREQUENCY DISTRIBUTION OF SUSPENDED SOLIDS IN NORTH SIDE MICROSTRAINER

FIGURE 15



capacity tests depicted in Figures 12 and 13. It can be seen that the secondary effluent from the North Side Plant is of high quality having an average (50 Percentile) S.S. of 11.5 mg/l and 87 percent of the values less than 20 mg/l. It also can be seen that the effluent S.S. from the unit is of high quality having an average (50 Percentile) S.S. of 3.2 mg/l with no value exceeding 12 mg/l.

In Figure 16 is presented a cummulative frequency distribution of the influent and effluent BOD from the unit. The average BOD (50 Percentile) of the influent to the unit was 10.0 mg/l while 85% of the values were less than 15.0 mg/l. Effluent BOD averaged (50 Percentile) 3.8 mg/l while no value exceeded 13.0 mg/l.

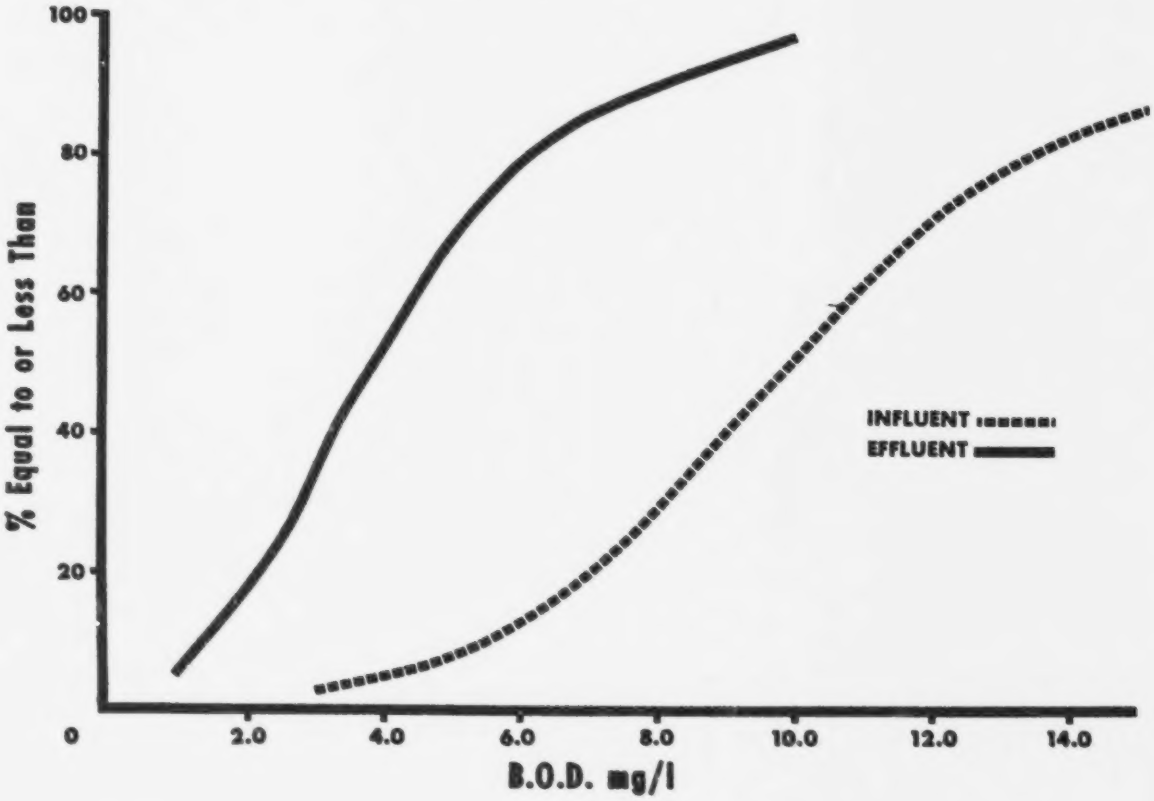
Backwash consumption for the unit it depicted in Figure 17. It can be seen that the backwash is a decreasing function of flow. This is because for the North Side unit, backwash flow remains fairly constant over the flow rates tested. Therefore, as flow through the unit increases, percent backwash decreases.

The backwash data indicates that for the average flow (7.6 mgd) processed by the unit during these tests, the unit required 4.6% backwash while at maximum flow (15 mgd) required about 3.0%.

THE METROPOLITAN SANITARY DISTRICT OF GREATER CHICAGO

CUMULATIVE FREQUENCY DISTRIBUTION OF B.O.D. CONCENTRATIONS IN NORTH SIDE MICROSTRAINER

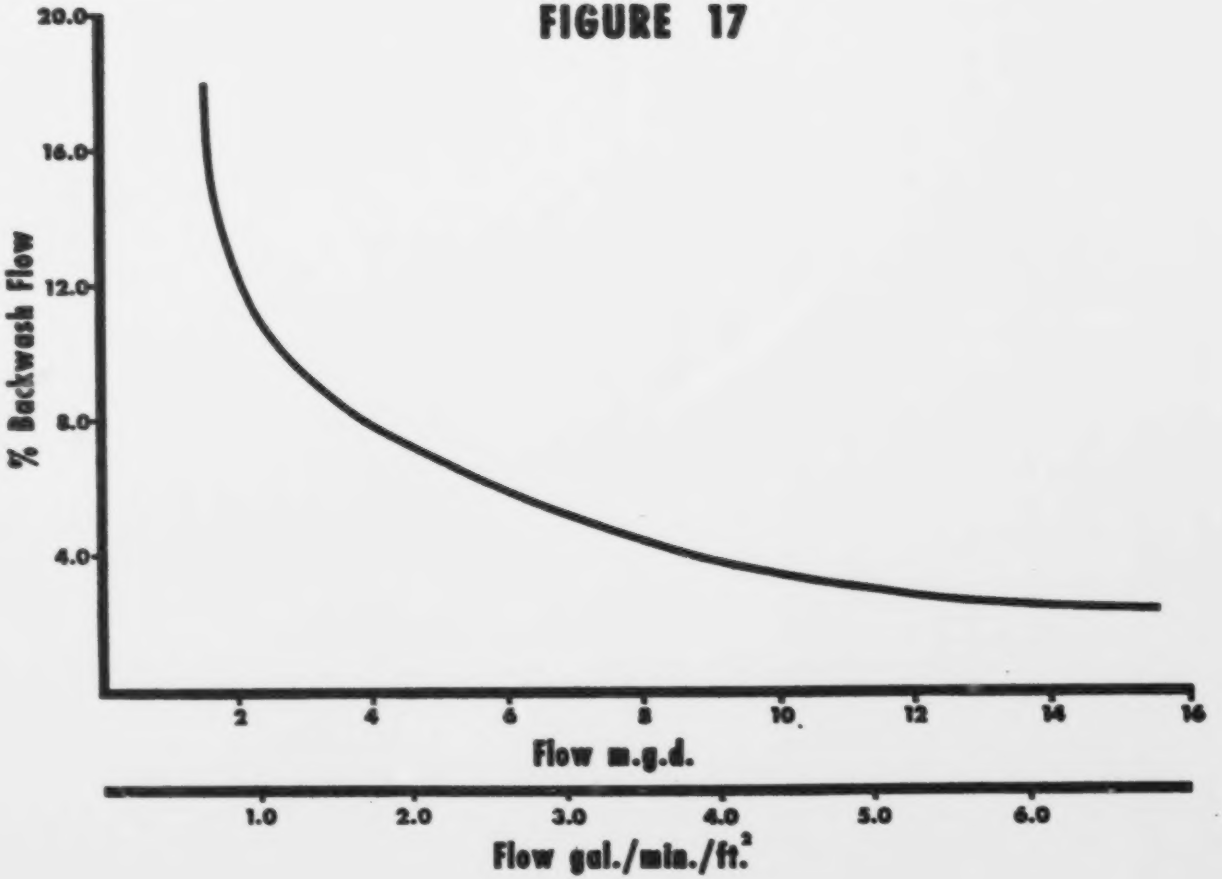
FIGURE 16



THE METROPOLITAN SANITARY DISTRICT OF GREATER CHICAGO

BACKWASH DATA FOR NORTH SIDE MICROSTRAINER

FIGURE 17



CONCLUSIONS

The results of the microstrainer tests at the North Side Plant did not prove as promising as first envisioned. Although effluent quality was satisfactory and well below present Illinois standards (Illinois standards require 30-day average BOD values of less than 10 mg/l and S.S. less than 12.0 mg/l), the flow capacity of the unit was well below that envisioned by the MSDGC.

The North Side unit has proven itself capable of taking flows on the average (50 Percentile) of about 7.6 mgd with corresponding solids loadings averaging (50 Percentile) about 0.5 lbs/ft^2/day. Effluent quality for the Mark O stainless steel fabric appears to be capable of meeting present Illinois S.S. and BOD standards. Backwash valume for the unit operating at an average flow capacity was about 4.6%.

SUMMARY COMPARISON OF TEST RESULTS OF FILTRATION DEVICES TESTED

In Table 4 is contained a summary of the operating and performance parameters for the filtration devices discussed in this paper. The following may be concluded from a comparison of these summarized results.

1. The North Side Microstrainer although capable of achieving the maximum flow capacity of the Hanover Microstrainer on a short-term basis, could not consistently achieve the maximum flow for the Hanover unit. This could indicate that microstrainer flow capacity is severly dependent on the type of effluent solids.

2. It appears that the microstraining devices cannot achieve the S.S. removal performance of sand filters.

3. It appears that batch type filters are capable of loadings over 1.2 lbs/ft^2/filter run.

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TABLE 4

COMPARISON OF OPERATING AND PERFORMANCE PARAMETERS OF THE FILTRATION DEVICES TESTED BY THE MSD

PARAMETER	HANOVER PARK MICROSTRAINER	NORTH SIDE MICROSTRAINER	HARDINGE SAND FILTER	DELAVAL SAND FILTER	NEPTUNE MICROFLOC SAND FILTER	GRAVER SAND FILTER
FLOW RATE (GPM/SQ FT)						
Maximum	6.6	6.3	6.0			
Range Tested	0-6.6	0-6.3	0-6.0	0-5.0	0-4.0	0-9.0
LOADING						
Maximum	0.88 **	1.0 **	0.65 **			
Range	0-0.88 **	0-1.0	0-0.65 **	0-1.24 **	0-1.53	0-1.9 *
BACKWASH (%)						
At Max. Flow	4.0	2.9	0.65			
Range	0-3.0	0-16	0-0.65	0-3.7	0-2.5	0-1.14
EFFLUENT BOD (MG/L)						
At Max. Flow	4.0 ■		2.3 ■			
Range	1-5	0-11	1-4	5-9	4.0-6.0	3.5-8.0
EFFLUENT S.S. (MG/L)						
At Max. Flow	5.0 ▲		5.0 ▲			
Range	1-10	0-12	1-9	5-7	4.0	5.9

■ For Influent BOD of 18 mg/l and Maximum Loading

▲ For Influent S.S. Equals 18 mg/l and Maximum Loading

* lbs/sq ft/filter Run

** lbs/sq ft/day

METRO CHICAGO

STUDIES ON NITRIFICATION

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THIRD U.S./JAPAN CONFERENCE ON SEWAGE TREATMENT TECHNOLOGY
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INTRODUCTION

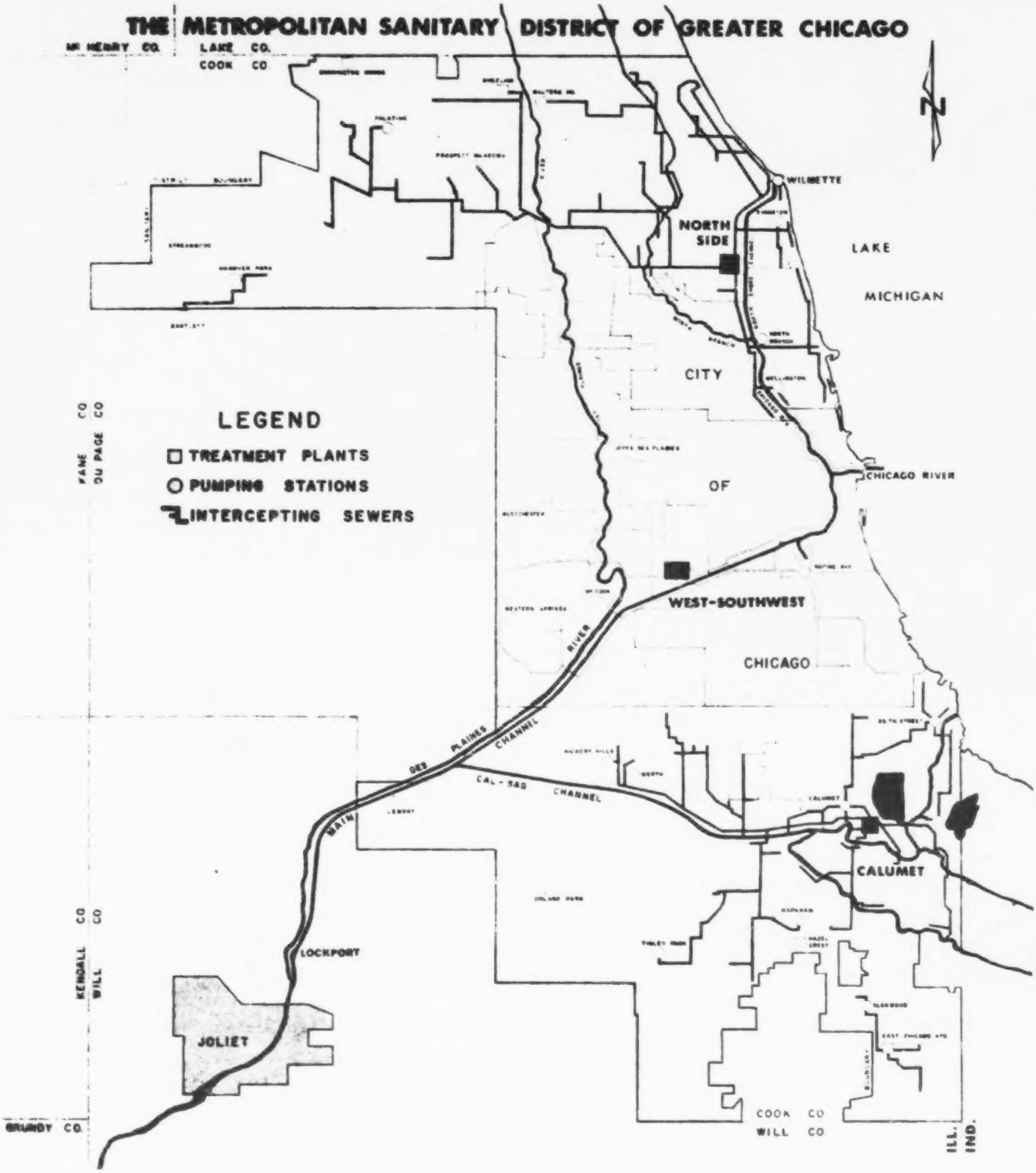
The Metropolitan Sanitary District of Greater Chicago (MSDGC) collects and treats more than one billion 250 million gallons of domestic and industrial waste each day. It has a capacity of 1,600,000,000 gallons a day and its growth and ungrading continues in order to keep pace with the needs and demands of the communities which it serves and to produce higher quality effluents in order to upgrade all water resources in the State of Illinois.

The District maintains a waterway system which consists of 85 miles of navigable waters including canals, channels and rivers (Figure 1). In addition, hundreds of miles of secondary tributaries and storm sewers also drain into the main waterway system. Because the District's three major sewage treatment plants and four small treatment plants discharge effluent to the drainage basin, more stringent criteria are being placed on the quality of effluent discharged. These criteria cover a wide spectra such as carbonaceous oxygen demand, toxic substances, nutrients and oils and greases. The MSDGC has directed considerable efforts toward the removal of nitrogen and phosphorus nutrients in order to develop effective and economical procedures.

It is now realized that besides it's role as an indicator of pollution, ammonia itself is a serious pollutant.(1,2) Ammonia nitrogen in waste treatment effluents has been held undesirable due to the following reasons:

DISTRICT MAP

FIGURE 1



1. It reacts with chlorine during the disinfection process and produces chloramines which are less effective as disinfectants than free chlorine and may subsequently cause an increase in the chlorine demand.

2. This constituent exerts a toxic effect on aquatic life at high concentrations. This has been demonstrated by investigators to be particularly toxic on fish life.

3. It exerts an oxygen demand on the receiving waterways and will subsequently, when oxidized to other nitrogenous forms act as a nutrient for undesirable algae forms.

The Illinois Pollution Control Board (IPCB) Rules and Regulations, Chapter 3, Water Pollutions require in Rule 406(3) that "No effluent from any source which discharges to the Illinois River, the Chicago River System, or Calumet River System, and whose untreated waste load is 50,000 or more population equivalents shall contain more than 2.5 mg/l of ammonia nitrogen as N during the months of April through October, or 4 mg/l at other times, after December 31, 1977." This rule will require that the three major treatment plants of the Metropolitan Sanitary District of Greater Chicago, namely the

North Side Sewage Treatment Works, the West-Southwest Sewage Treatment Works, and the Calumet Sewage Treatment Works remove ammonia nitrogen from their effluents after December 31, 1977.

To this end the MSDGC has undertaken pilot process studies to evaluate alternate procedures for removing ammonia nitrogen from its secondary treatment effluents.

After serious consideration of the physical-chemical and biological processes available for ammonia removal, it was decided that secondary or unresolved problems made the physical-chemical processes unfeasible at this time. (4,5) Therefore, the MSDGC concentrated its efforts on the biological processes for ammonia removal.

This paper will present some of the data collected on several of the biological treatment projects conducted from 1969 through 1973. These data will be discussed in the following order:

1. Two-Stage Nitrification at Hazelcrest, Illinois
11. Calumet Nitrification Pilot Plant
111. Nitrification at Lemont, Illinois - Single Stage Extended Aeration
1V. Nitrification of a High Ammonia Content Sludge Supernatant by Biological Processes

Two-Stage Nitrification at Hazelcrest, Illinois

The 1.2 mgd design capacity treatment facilities at Hazelcrest, Illinois, were converted from two parallel activated sludge systems to a two-stage biological nitrification process. The total capacity of the nitrification plant was approximately 0.6 mgd.

In April of 1969, the Hazelcrest facility began operating as a two-stage nitrification process. However, it was not until June of the same year that the system attained a satisfactory level of nitrification. This level of nitrification continued through the month of December 1969. During this period of operation, the detention time ranged from 4.5 to 3.6 hours for the first and second stage reactors. The first and second stage clarifier detention times ranged from 1.95 to 1.56 hours and 3.22 to 4.03 hours respectively. These detention times were calculated on an average flow of 0.8 and 1.0 mgd which were the range of flows generally experienced by the plant.

Results

The summary data of the Hazelcrest project will be presented in two phases. Phase I will show data for the initial start-up and a period of successful operation. Phase II will show data after the nitrification process was interrupted and steps were taken to re-establish it.

Phase I - Initial Testing Period (April 20 - July 30, 1969

As in any start-up period, there were fluctuations in plant operation. Once the plant reached a point of steady state operation, mixed liquor suspended solids in the first stage reactor had to be maintained between 1800 and 2000 mg/l. If the mixed liquor solids were allowed to exceed 2000 mg/l, excess solids were carried over into the second stage.

On May 1, solids wasting from the second stage was discontinued to build up the mixed liquor solids. By following this procedure mixed liquor solids concentrations were maintained between 2800 and 3000 mg/l. It was only necessary to waste occasionally to prevent second stage effluent deterioration.

The data obtained during this period of study will be presented in the next four figures. It can be seen from the nitrate - nitrite and ammonia curves (Figure 2) that the plant approached a satisfactory point of nitrification around June 23 and continued to nitrify at this level through July 30. The average raw NH_3-N concentration of 12 mg/l was reduced to an average concentration of 1.5 mg/l in the second stage effluent. The nitrate-nitrite concentration ranged from 10.0 to 17.0 mg/l. The temperature ranged from 50°F to 64°F during this phase of study.

As can be seen from Figures 3 and 4, the second stage reactor operated with an average loading of 0.10 lb BOD/lb

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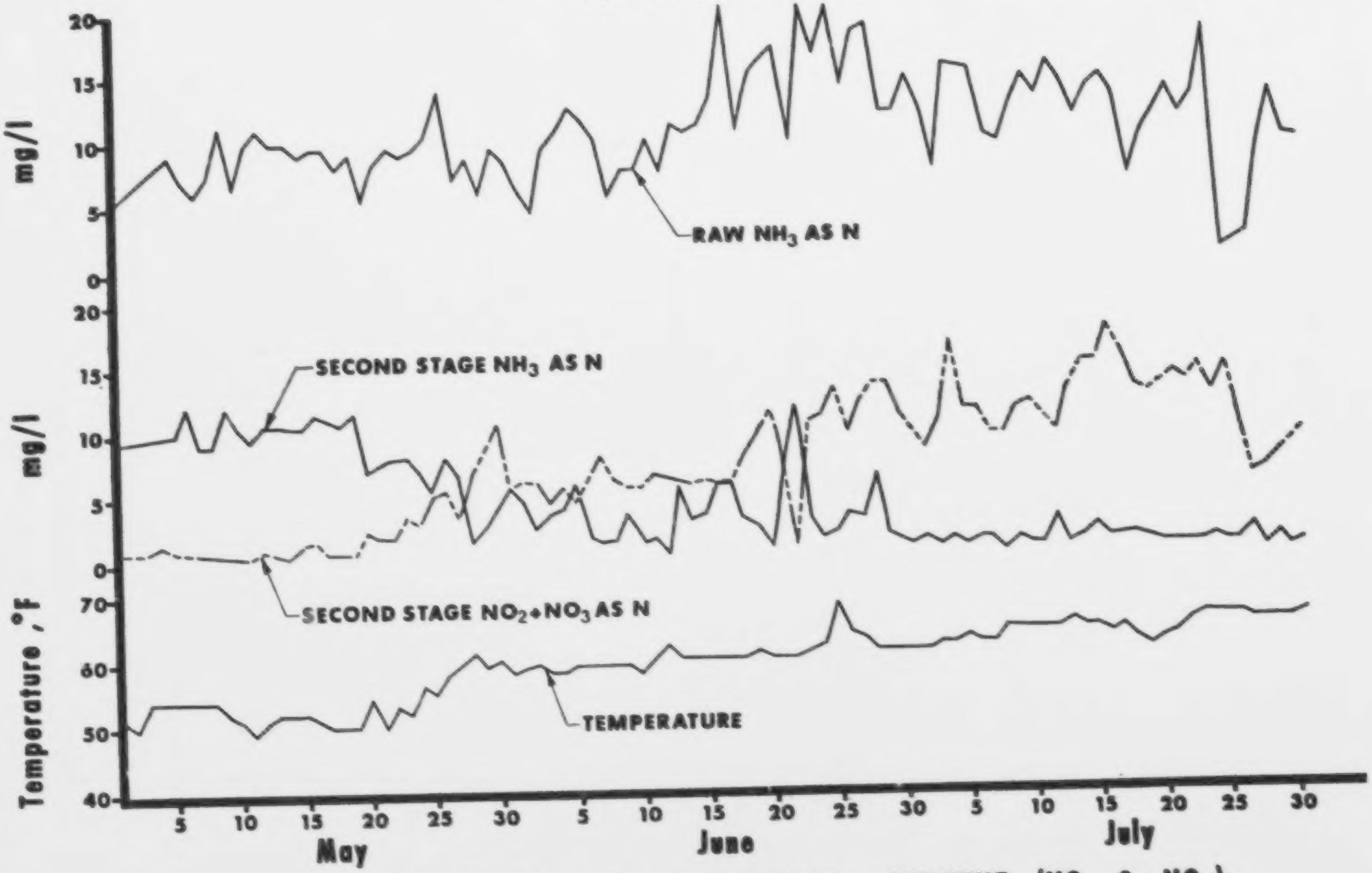
FIGURE 2



TIME SERIES PLOT OF RAW SEWAGE AMMONIA, EFFLUENT (NO$_2$ & NO$_3$) AND EFFLUENT AMMONIA AND RAW SEWAGE TEMPERATURE AT HAZELCREST

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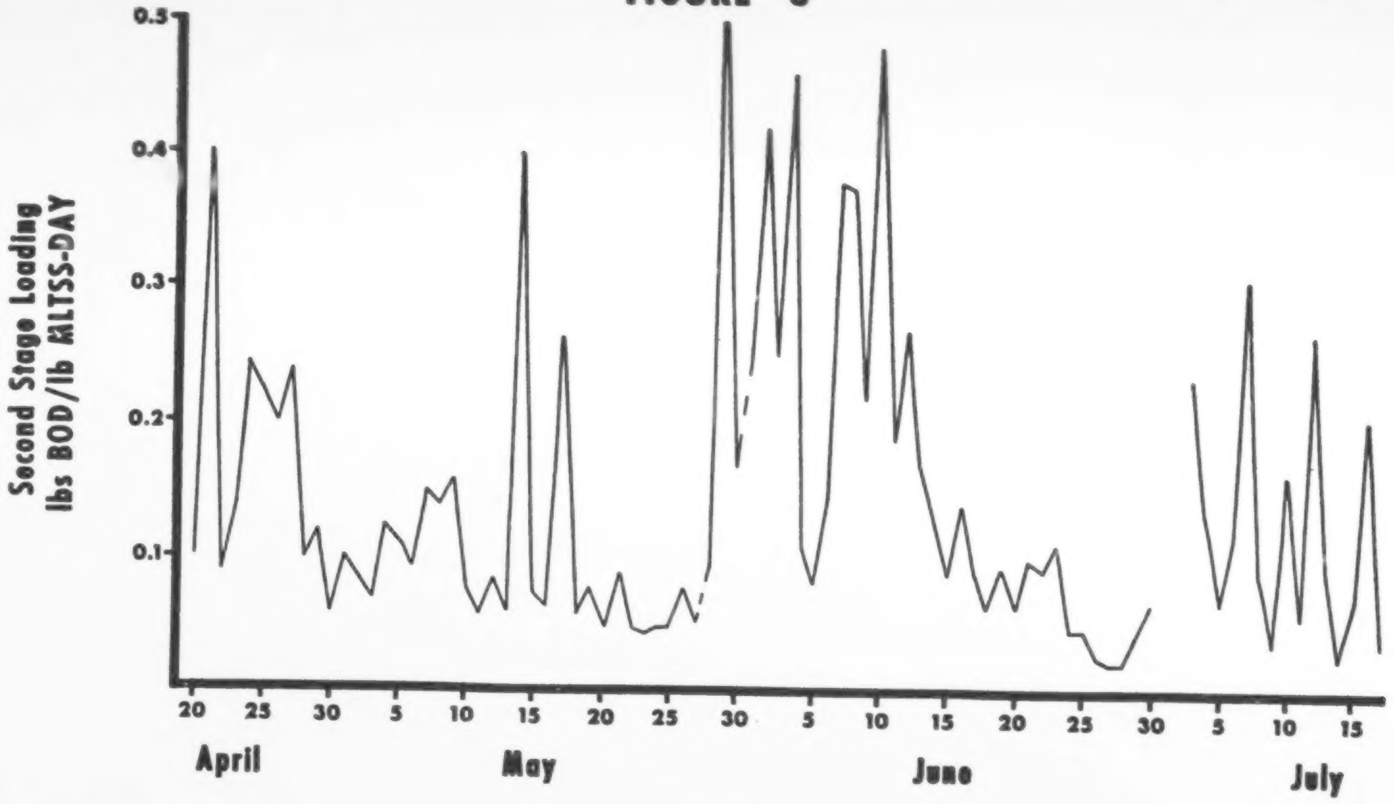
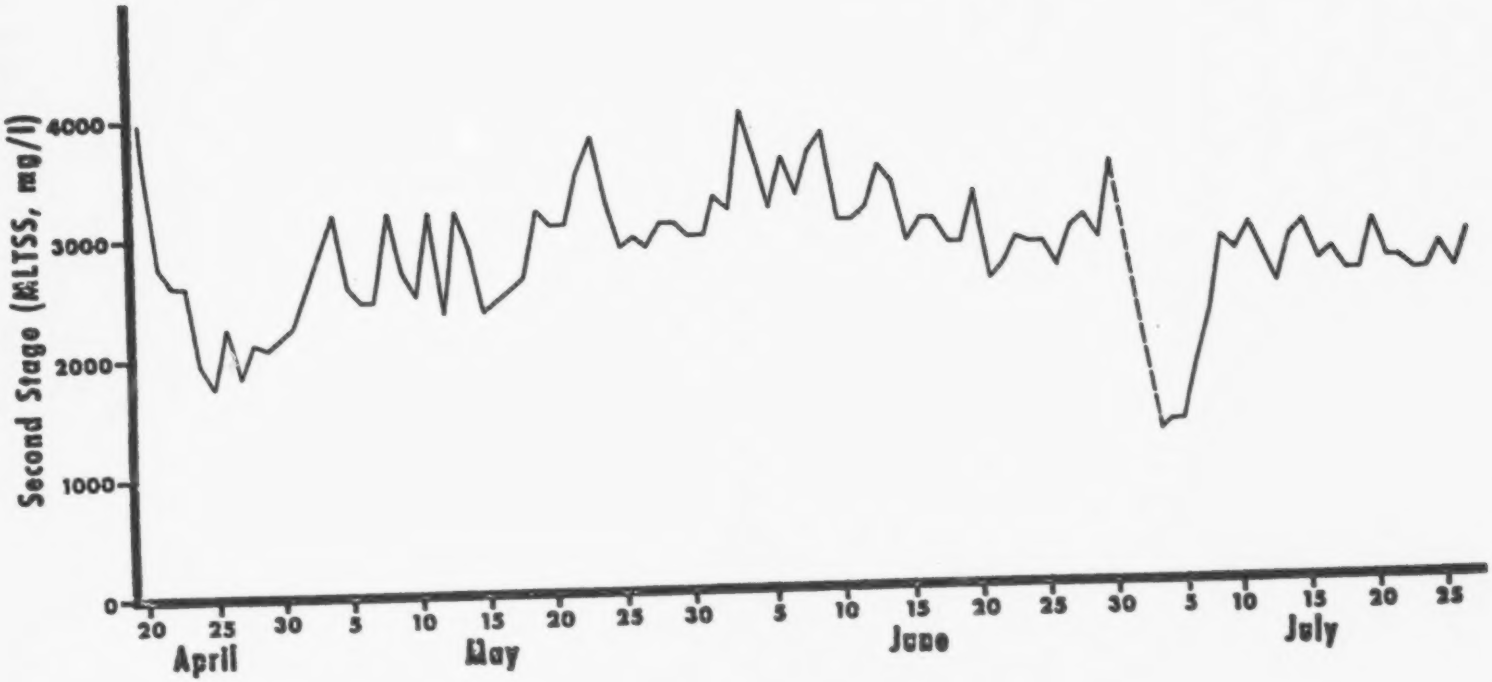
FIGURE 3



TIME SERIES PLOT SECOND STAGE B.O.D. LOADING AT HAZELCREST

METROPOLITAN SANITARY DISTRICT OF GREATER CHICAGO

FIGURE 4



TIME SERIES PLOT OF SECOND STAGE MLTSS AT HAZELCREST

MLTSS/day and a mixed liquor suspended solids range of 2500 to 3500 mg/l from April 20 to July 30, 1969.

Figure 5 presents data on the NH_3-N loading to the second stage. The loading was generally less than 0.04 lbs NH_3/lb MLTSS/day. The plant continued to nitrify at about 90% efficiency with NH_3-N less than 1.5 mg/l through the month of October.

To select or choose factors responsible for achieving the degree of nitrification observed is difficult but one may consider several factors. A comparison of the data presented in the graphs for the period of satisfactory and unsatisfactory nitrification is of interest. The organic loading to the second stage was somewhat lower during satisfactory nitrification. Also, sewage temperature appears to exert a significant effect on nitrification. From April 20 through May 21, the temperature was 50°F at which time there was no nitrification. There was some nitrification when the temperature increased to 59and 60°F. The temperature remained at this level for 30 days and the plant was nitrifying at 50% efficiency. When the temperature increased to 63 and 64°F the plant began to produce complete nitrification.

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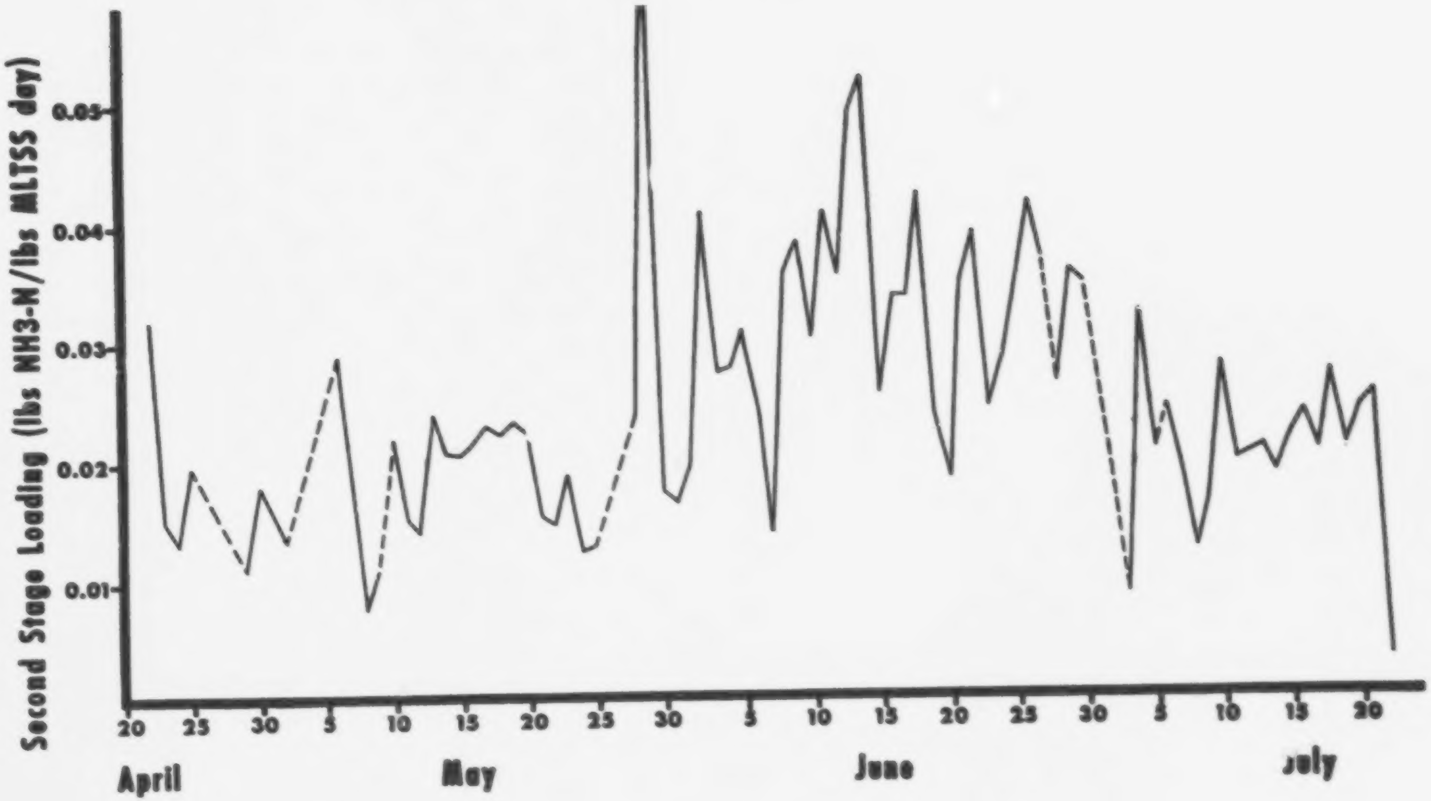
FIGURE 5



TIME SERIES PLOT OF SECOND STAGE AMMONIA LOADING
AT HAZELCREST

Phase II - December 15, - March 24, 1970

As previously noted, the Hazelcrest Plant produced a well nitrified effluent during the months of November and December. However, about January 1, 1970 a decrease in second-stage nitrate levels and a corresponding increase in the NH_3-N level occurred. This decrease in plant efficiency is depicted in Figure 6. A clear indication of the exact cause of the cessation of nitrification is not evident but certainly several factors must be considered to have significantly contributed to the effect.

The organic loading to the second stages (Figure 7) was high, averaging about 0.3 lb BOD/lb MLTSS/day. As indicated in Phase I, a more reasonable figure for the second stage loading would be 0.1 lb BOD/lb MLTSS/day.

Another factor contributing to the lack of nitrification was the relatively low raw sewage temperatures that occurred January 1, 1970, and thereafter (Figure 8). This average temperature of 47°F was much lower than the temperature experienced in the spring of 1969, at which time partial nitrification was obtained.

It was also possible (Figure 9) that the relatively large fluctuations in MLTSS did not contribute favorably to the nitrification process. The inability to maintain a steady-state operation with respect to the above parameters was certainly a

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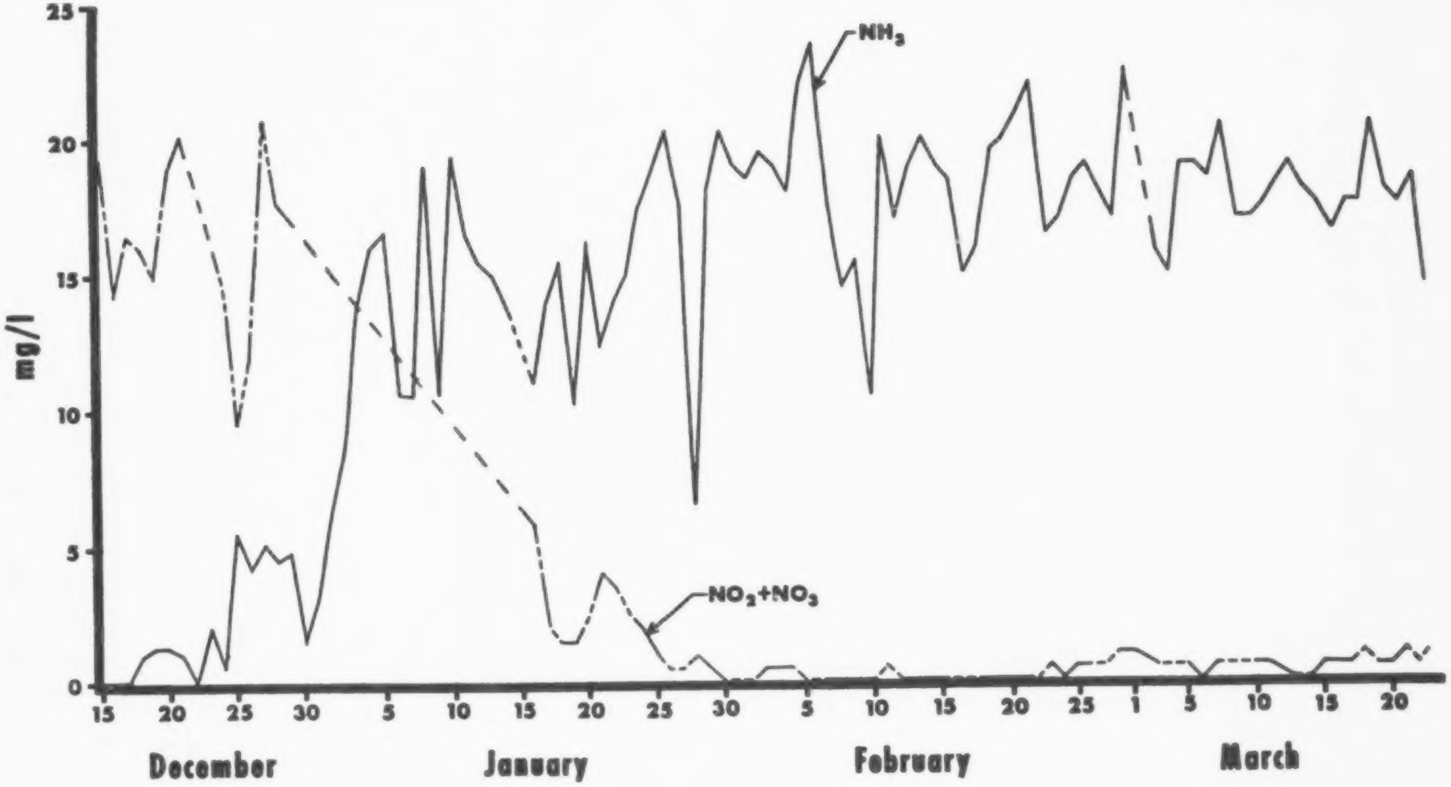
FIGURE 6



TIME SERIES PLOT SECOND STAGE EFFLUENT AMMONIA AND (NO_2+NO_3) AT HAZELCREST

]

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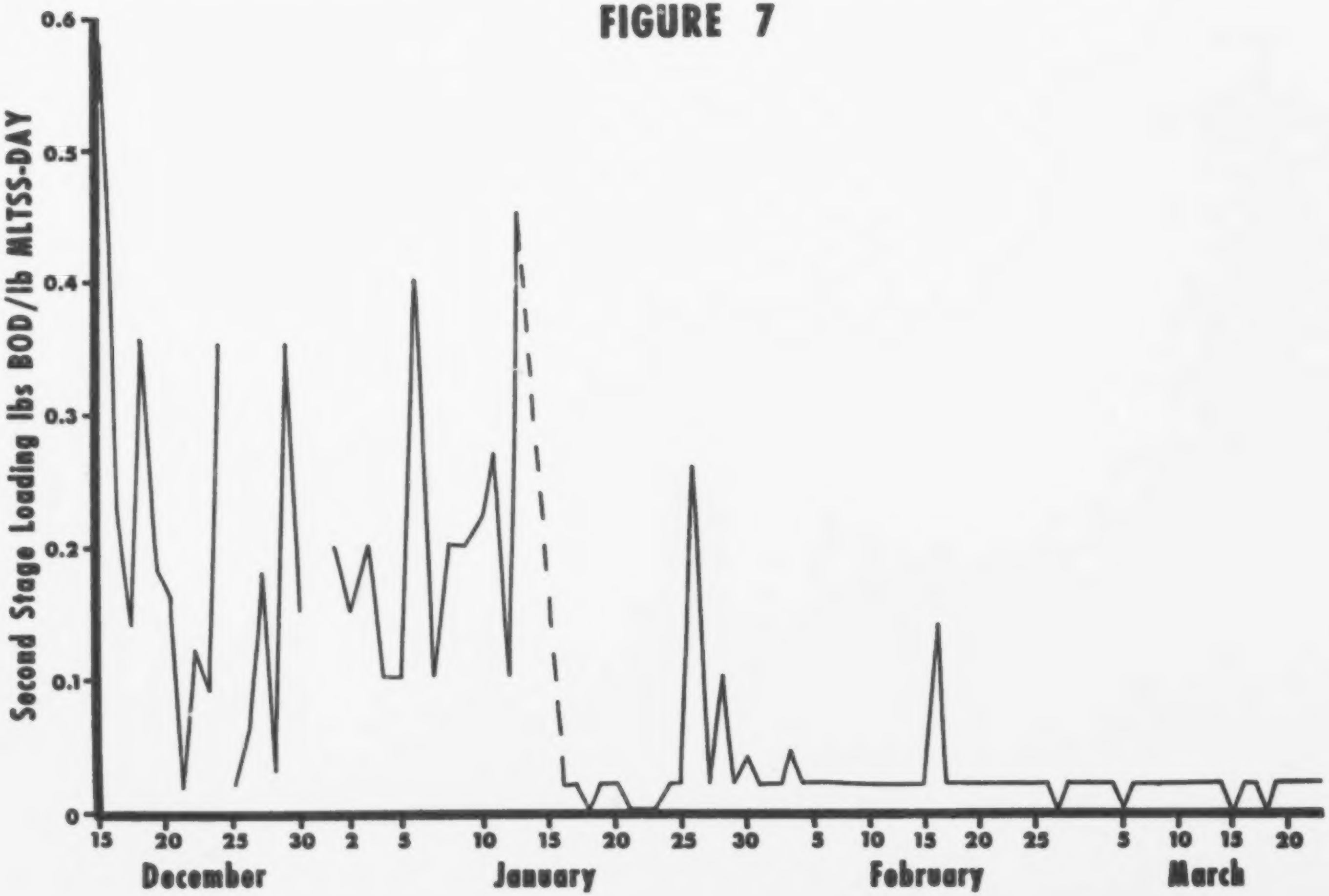
FIGURE 7



TIME SERIES PLOT SECOND STAGE B.O.D. LOADING AT HAZELCREST

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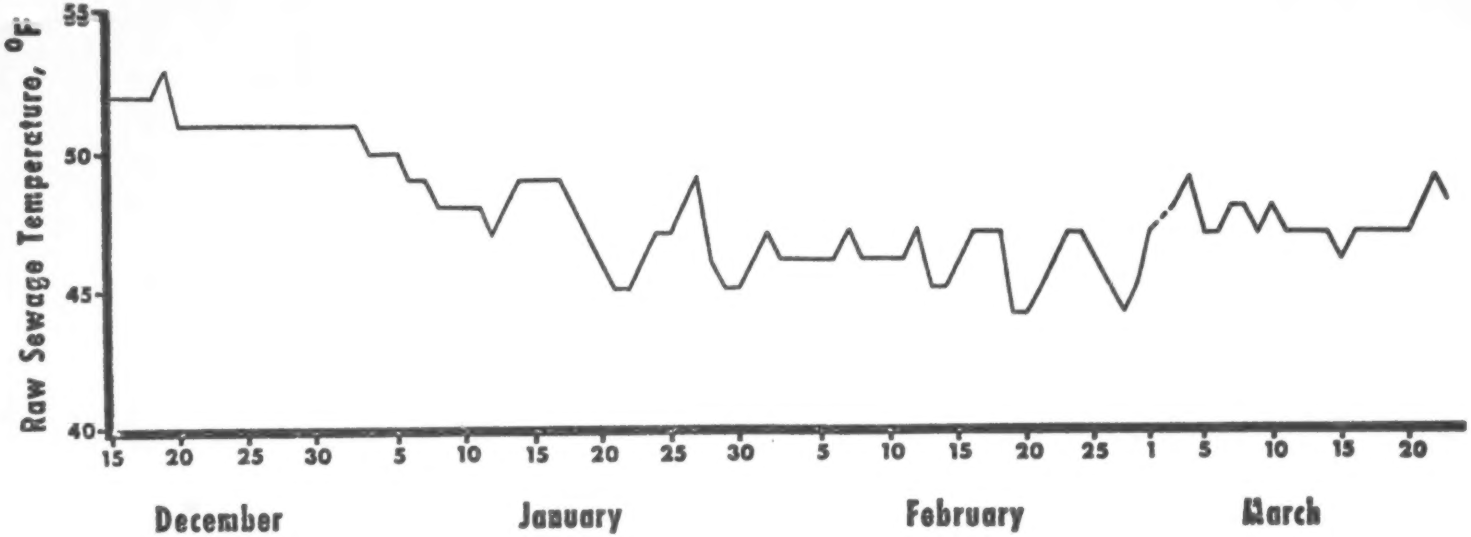
FIGURE 8



TIME SERIES PLOT RAW SEWAGE TEMP. AT HAZELCREST ILL.

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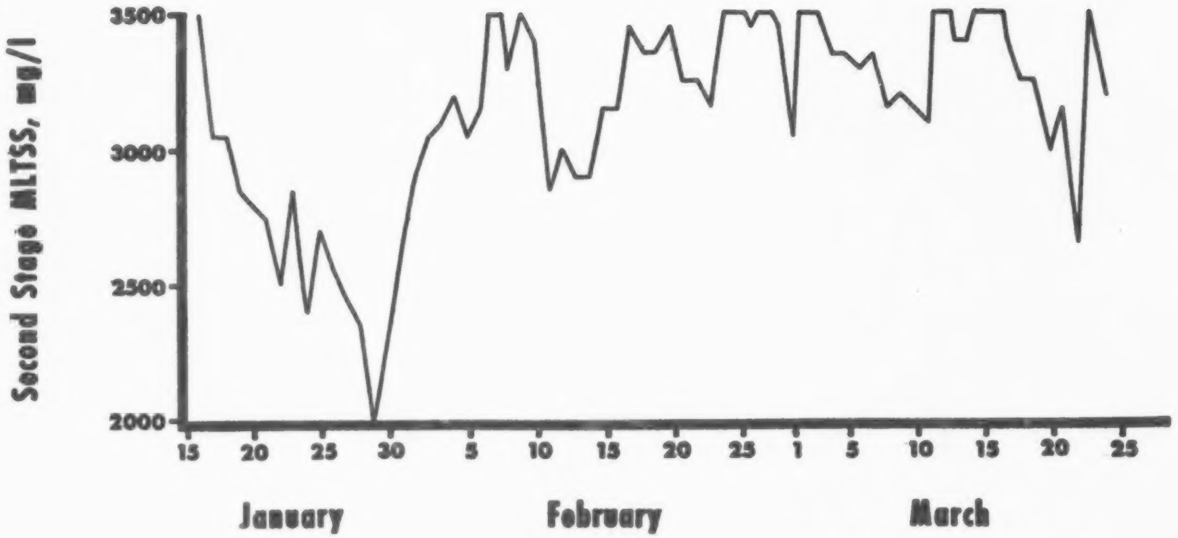
FIGURE 9



TIME SERIES PLOT SECOND STAGE MLTSS AT HAZELCREST detriment to the system.

Due to the absence of flexibility in the operation of the plant, it was difficult to determine precisely the total or direct effect of the various parameters believed to exert adverse changes on the nitrification process. However, the following conclusions were drawn from the MSDGC experiences with this study.

1. A two-stage activated sludge process can produce an effluent high in nitrate - nitrite and low in ammonia concentrations at a detention time of 3.0 to 5.0 hours in the second stage.

2. It is necessary for the first-stage system to perform efficiently in organic and suspended solids removal to prevent detrimental carry-over of these materials into the second stage.

3. Temperature effects appear to be a major factor in initiating and maintaining nitrification.

4. A highly nitrified effluent can be produced when the NH_3 loading is less than 0.04 lbs/MLVSS/day to the second stage.

Calumet Nitrification Pilot Plant

Introduction

The Calumet Sewage Treatment Plant, located in the southeastern section of the City of Chicago and serving this general portion of the city and Cook County, was placed into operation in 1935. The present plant receives combined domestic and industrial sewage from an area of approximately 270 square miles. The plant's sewage processing facilities include preliminary and secondary treatment, anaerobic sludge digestion, and contact chlorination.

There is a wide range of industries operating within the Calumet Treatment Plant service area, such as steel mills, coking plants, paint and chemical manufacturers, and grain and transport companies. These and other regional industries contributed an estimated 60 - 70 mgd of the average 199 mgd of sewage volume which the Calumet Plant processed in 1972. Table 1 lists some of the pertinent characteristics of the Calumet raw sewage obtained during the six months in which the Calumet Nitrification Pilot Plant was operated.

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TABLE 1

CALUMET NITRIFICATION PILOT PLANT
CALUMET RAW SEWAGE CHARACTERISTICS
MONTHLY AVERAGES DURING PILOT STUDY

	TKN	NH_3-N	pH	TSS	VSS	BOD	COD
NOV. 1972	24.1	15.6	7.5	194	121	106	245
DEC.	29.9	17.8	7.5	265	157	149	330
JAN. 1973	28.3	18.0	7.4	178	113	136	287
FEB.	32.4	23.7	7.4	174	107	148	312
MAR.	24.4	17.6	7.5	192	108	130	262
APR.	22.8	15.7	7.6	231	116	113	234
AVERAGES	27.0	18.1		206	120	130	278

Results expressed in mg/l, except for pH.

Results

From November 1, 1972, through April 24, 1973, a two-stage bench-scale nitrification feasibility study was conducted at Battery C of the Calumet Sewage Treatment Plant. The final effluent from Battery C was utilized as the influent feed to a 20.6 gallon, compartmentalized pilot reactor which constituted the second stage of a two-stage biological treatment process. A schematic of the pilot plant is shown in Figure 10. Within three days of start-up of the pilot plant, significant nitrification was observed and was maintained at approximately 90 per cent ammonia nitrogen removal until the completion of the project.

The data collected for the nearly six months of operations has been summarized as follows:

1. Substrate Loading and Substrate Removal

   The influent feed ammonia nitrogen averaged 17.5 mg/l, with the weekly averages ranging from 10.1 to 24.9 mg/l; this included a supplemental 10 mg/l NH_4Cl dosing to the reactor which was initiated during the latter phase of the study. The average ammonia nitrogen reactor loading rate was 0.126 lbs NH_3-N/lb MLVSS/day; weekly averages ranged from 0.041 to 0.273 lbs NH_3-N/lb MLVSS/day. Effluent ammonia nitrogen concentrations averaged 2.1 mg/l, an average of 88 per cent removal. The average ammonia nitrogen removal rate was 0.109 lbs NH_3-N/lb MLVSS/day.

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CALUMET NITRIFICATION PILOT PLANT

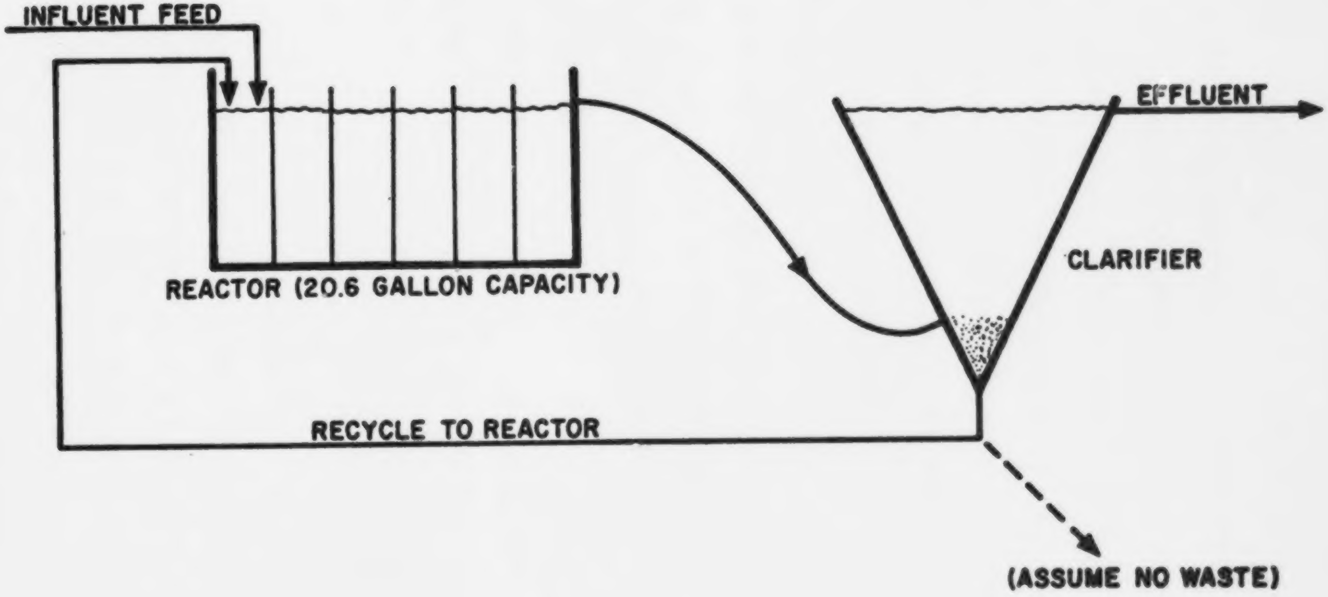

FIGURE 10

SCHEMATIC OF 2ND-STAGE BIOLOGICAL NITRIFICATION SYSTEM

Figure 11 details the variability in ammonia nitrogen for influent and effluent streams and the resultant % removal for the initial five days of start-up, and the subsequent 25 weekly averages. The effluent ammonia nitrogen observed for weeks 15, 16, and 17 were high, averaging 4.6 mg/l, due to insufficient mixing in the aeration tank which resulted in inadequate oxygen transfer. However, as indicated, the effluent quality was greatly improved by the 18th week, following correction of the mixing problem. Again, during the period between the 21st week and the 23rd week of operations, operational difficulties (which in this case was the loss of the nitrifying population as a consequence of the reactor overflowing) resulted in high ammonia nitrogen bleed-throughs to the effluent. The residual effluent ammonia nitrogen obtained for this period averaged 3.4 mg/l. As was previously stated, effluent NH_3-N averaged only 2.1 mg/l over the full 25 weeks of the study.

2. Effect of Change in Detention Time

The nitrification reactor was successfully operated at detention times of 4.0, 3.0 and 2.0 hours, based on influent flow rate only, for periods of 82, 42 and 51 days, respectively. Throughout the study recycle sludge

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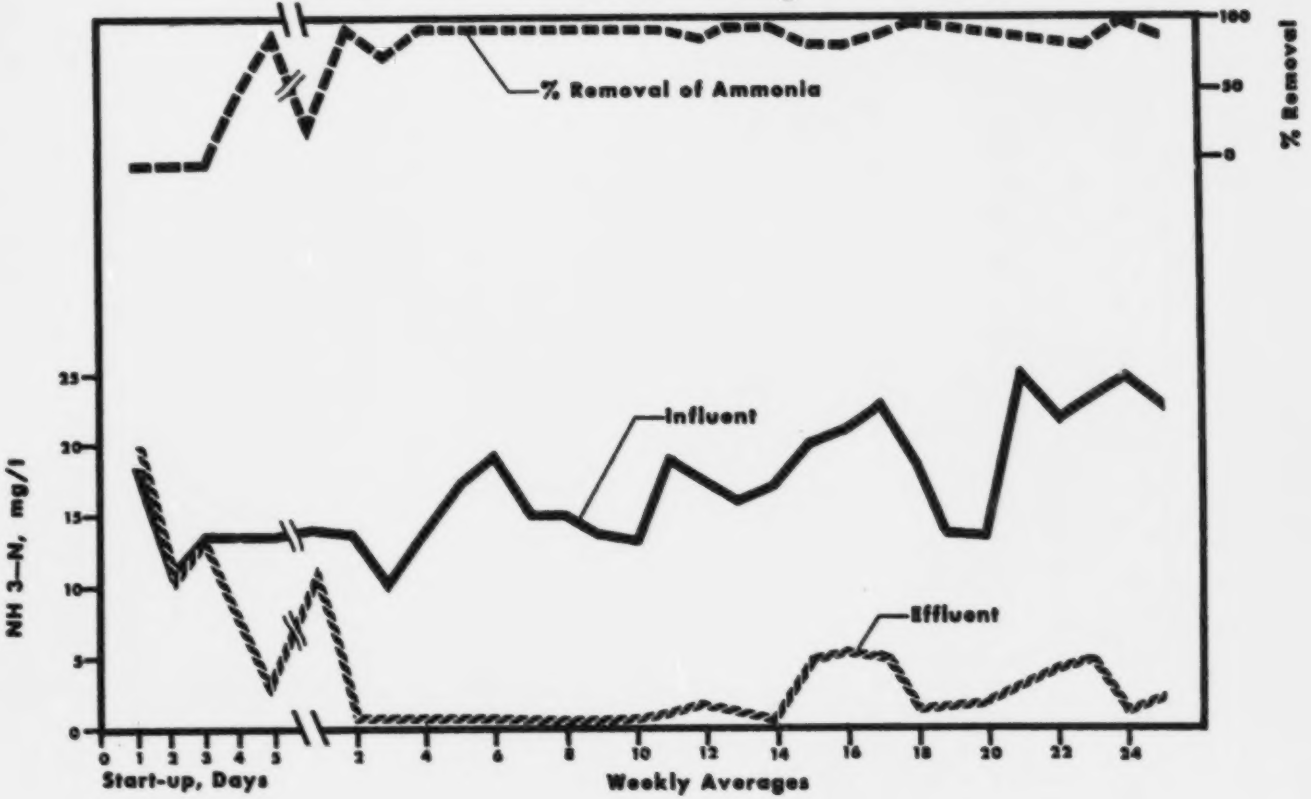
FIGURE 11

CALUMET NITRIFICATION PILOT PLANT

Pilot Plant Influent and Effluent NH_3-N Concentrations



was pumped to the head of the reactor at rates of either 50% or 75% of the influent flow. Excluding the initial start-up week, the average effluent ammonia nitrogen was 0.9, 2.8 and 2.0 mg/l, respectively, for the detention times investigated.

3. Solids Retention Time, SRT

For the first 11 to 12 weeks of this study, efforts were not directed towards maintaining a sludge age of any particular number of days. However, for the remaining weeks of the study, attempts were made to control the sludge age at an average of at least ten days. This was done by periodically adding settled return sludge from Battery C to the nitrification reactor whenever effluent suspended solids losses and/or accidental sludge wasting resulted in MLVSS levels of approximately 1000 mg/l or less. The average SRT obtained during this latter phase of operations was nine days at an average reactor temperature of 15°C.

4. BOD and Suspended Solids Considerations

Influent BOD and TSS to the second-stage reactor averaged 18 mg/l and 13 mg/l, respectively, whereas effluent BOD and TSS averaged, respectively, 29 mg/l and 26 mg/l. Typically, BOD and suspended solids removal through the nitrification unit was not observed once nitrification commenced. Figure 12 indicates the frequency distribution of effluent TSS which was obtained for each of the two major clarifier overflow rates at which the clarifier was operated, namely 287 gal/sf/day and 382 gal/sf/day. As shown, the higher effluent TSS were more often observed at the higher clarifier overflow rate.

5. Maintenance of Mixed Liquor Volatile Suspended Solids

Weekly averages of mixed liquor suspended solids ranged from 1151 mg/l to 3642 mg/l, averaging 2115 mg/l of which an average 1359 mg/l (64.2%) was MLVSS. It was necessary to periodically add settled return sludge from Battery C to the nitrification unit in order to maintain a level of 1000 - 1500 mg/l MLVSS. The addition of return sludge to the reactor occasionally resulted in a transient upset in nitrification efficiency. This was adjudged to be due to dilution of the nitrifying population and was considerably lessened by a slow addition of the solids to the reactor.

6. Temperature Observations

Since there were no provisions for controlling the

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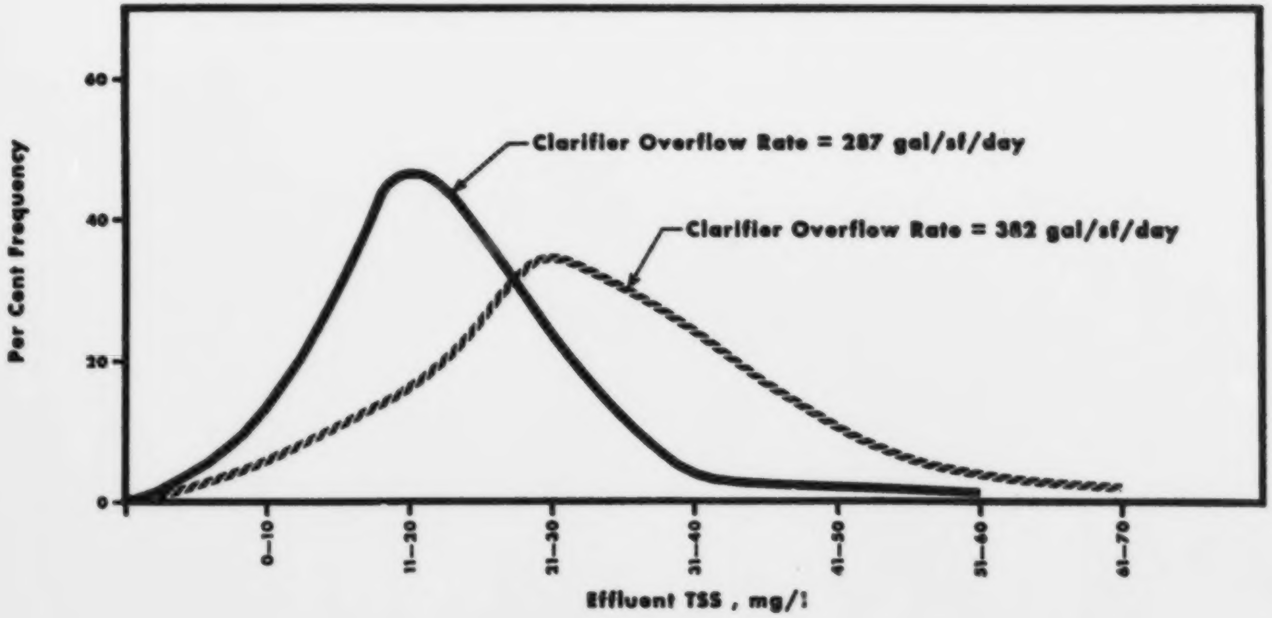
CALUMET NITRIFICATION PILOT PLANT

FIGURE 12



Frequency Distribution of Effluent TSS from the Calumet Nitrification Pilot Plant temperature of the nitrification unit, aeration temperatures varied with changes in feed and ambient temperatures. Average weekly reactor temperatures ranged from 22°C to 13°C over the six months involved in the study.

7. pH and Alkalinity Observations

pH control did not present any operating problems with the Calumet Nitrification Pilot Plant. The influent pH ranged from 7.4 to 7.7, and effluent pH ranged from 7.3 to 7.8 over the 25 weeks of the study. Sawyer, (Ref. 6) from his studies at Marlborough, Mass., recommended a mixed liquor pH range of 7.6 to 7.8 in order to allow carbon dioxide to escape to the atmosphere. This pH range would still be sufficiently close to the theoretical maximum pH of 8.4 to insure a high nitrification rate. The pilot plant mixed liquor pH's ranged from 6.9 to 7.7, and nitrification was apparently not attentuated at these pH levels.

Effluent alkalinity concentrations ranged from 60 to 650 mg/l, averaging 174 mg/l. The average alkalinity consumption was 5.2 lbs alkalinity (as $CaCO_3$)/lb NH_3-N oxidized and ranged from 1.2 to 7.8 lbs alk/lb NH_3-N oxidized.

8. Trace Metal Effects

Two aspects of high trace metal concentrations were considered in this study: (a) the effects of influent concentrations of trace metals on the nitrification process, and (b) the extent and effect of trace metal accumulation in the return sludge. Of the 14 metals monitored, none were found to be significantly high in either the influent or effluent streams. Typically, influent and effluent concentrations of metals were about the same; this is reflected in Figures 13 and 14 for two of the metals, Al and Cr. There were no occasions when upsets in nitrification efficiency were attributable to the concentrations of trace metals observed in the influent stream.

Table 2 lists the average return sludge trace metal concentrations for Calumet Battery C and also for the nitrification pilot plant. As shown, some of the pilot plant trace metal values exceeded those of Battery C, whereas others were less. This merely reflected the variability in trace metal concentrations that was observed for the pilot plant return sludge. This variability was due to periodic sludge losses through accidental sludge wastage and high effluent suspended solids, and sludge additions which were required to replenish those losses. Thus, as assessment of trace

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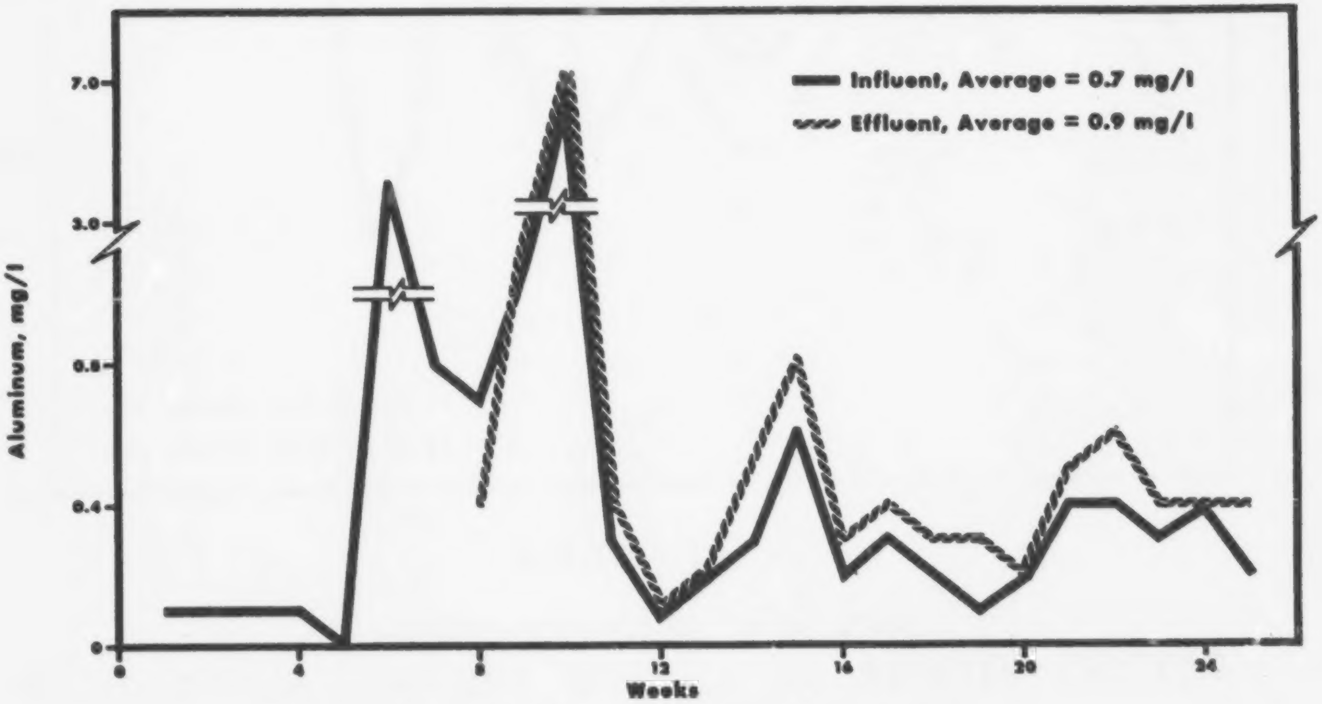
CALUMET NITRIFICATION PILOT PLANT

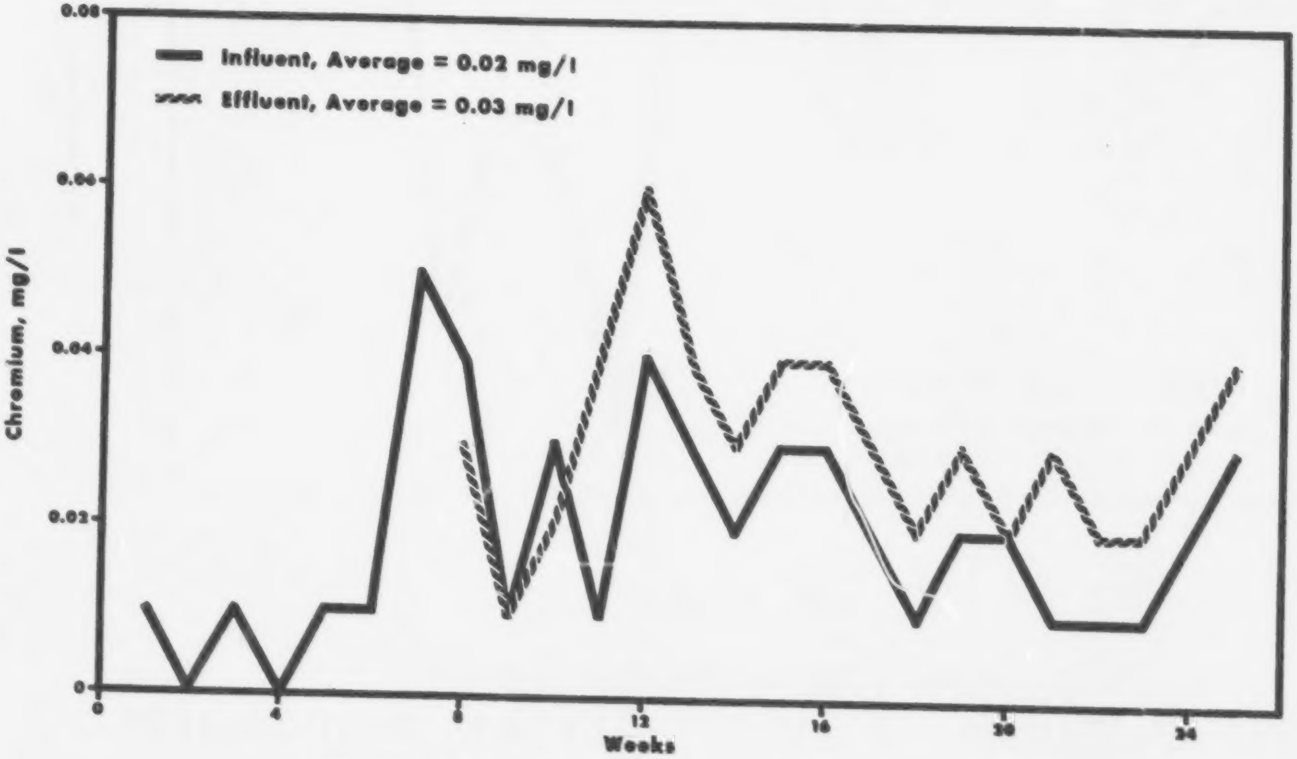
FIGURE 13



Weekly Variations in Aluminum Concentration

METROPOLITAN SANITARY DISTRICT OF GREATER CHICAGO
CALUMET NITRIFICATION PILOT PLANT

FIGURE 14



Weekly Variations in Chromium Concentration

METROPOLITAN SANITARY DISTRICT OF GREATER CHICAGO

CALUMET NITRIFICATION PILOT PLANT

TABLE 2

RETURN SLUDGE TRACE METAL ANALYSES & AVERAGES

(Results Expressed in mg/l except Hg. Hg in ug/g)

	CALUMET BATTERY 'C'	CALUMET PILOT PLANT
Zn	2.28	1.86
Cd	0.070	0.058
Cu	0.148	0.135
Cr	0.493	0.531
Fe	13.8	14.7
Ni	0.062	0.087
Pb	0.504	0.431
K	3.95	3.41
Na	11.0	19.4
Ca	43.2	37.9
Mg	13.4	13.5
Mn	0.256	1.121
Al	4.23	4.37
Hg	1.34	2.22

NOTE: All Values Calculated on a dry weight basis.

metal accumulation in the return sludge could not be properly evaluated.

9. Dissolved Oxygen Considerations

The amount of air supplied to the reactor was determined in great measure by the need to keep the MLSS in suspension via the bubbling actions of the compressed air. Thus, although the air input was measured, as was also the reactor dissolved oxygen, the values obtained were not reflective of what would be expected in a full-scale plant.

Conclusions

The following conclusions were derived from the results obtained by operating the Calumet Nitrification Pilot Plant:

1. The Calumet Treatment Plant, with a second-stage nitrification system, could reasonably be expected to produce a nitrified effluent with an ammonia nitrogen residual of less than 2.5 mg/l under the following conditions:

   influent NH_3-N, 10 - 25 mg/l
   SRT, at least 10 days
   minimum operating temperature, 12 - 15°C

   It must be emphasized that the lowest average temperature of 12°C recorded in the pilot plant aeration tank was approximately 2 - 3°C higher than that observed in the full-scale aeration tanks. Therefore, since a lower second-stage aeration temperature would be expected in an on-line plant, cold temperature nitrification studies on the Calumet sewage will be undertaken.

2. The average alkalinity consumption of 5.2 lbs alkalinity/lb NH_3-N oxidized that was required in the nitrification of Calumet Battery C effluent correlates reasonably well with the theoretical demand of 7.2 lbs/lb NH_3-N oxidized. Since Battery C effluent supplied this demand

without resulting pH depressions through the nitrification unit, it would be expected that future nitrification of Calumet sewage would not entail supplementation of the treatment stream alkalinity. However, in alkalinity deficient or high ammonia content wastewaters which are undergoing biological nitrification, it is expected that alkalinity additions will be required.

3. The preceeding first-stage sludge of the carbonaceous system apparently reduced influent concentrations of trace metals to non-inhibitory levels. However, the effects of trace metals upon nitrification in a single-stage system, under the conditions listed, are not known.

Nitrification at Lemont, Illinois - Extended Aeration

The Lemont Water Reclamation Plant (WRP) treats sewage primarily from the Village of Lemont, a town of about 5000 located in southwestern Cook County. The plant was designed in 1969 and completed in mid 1972. The schematic (Figure 15) shows the treatment processes employed at the plant. A brief description of plant facilities is as follows:

Design capacity	1.2 mgd
Maximum flow receiving complete treatment	3.0 mgd

A. Grit and screening

B. Primary settling
Two 45 ft. diameter tanks
Detention time at 1.2 mgd - 4.5 hr.
Surface settling rate - 380 gal/ft^2/day

C. Aeration tanks
Two single pass tanks 120 ft by 25 ft, 13.5 depth
Detention time at 1.2 mgd - 12 hr
Diffused aeration with 6000 cfm blower capacity
100% sludge return capacity

D. Final settling tanks
Two 36 ft. diameter center feed tanks
Detention time at 2.4 mgd mixed liquor flow - 3.7 hours
Surface settling rate at 2.4 mgd mixed liquor flow - 600 gal/ft/day

E. Microstrainers
Two 10 ft by 10 ft Zurn microstrainers
Design flow (each) - 1.2 mgd
Maximum flow (each) - 1.7 mgd
Screen size - 23 microns

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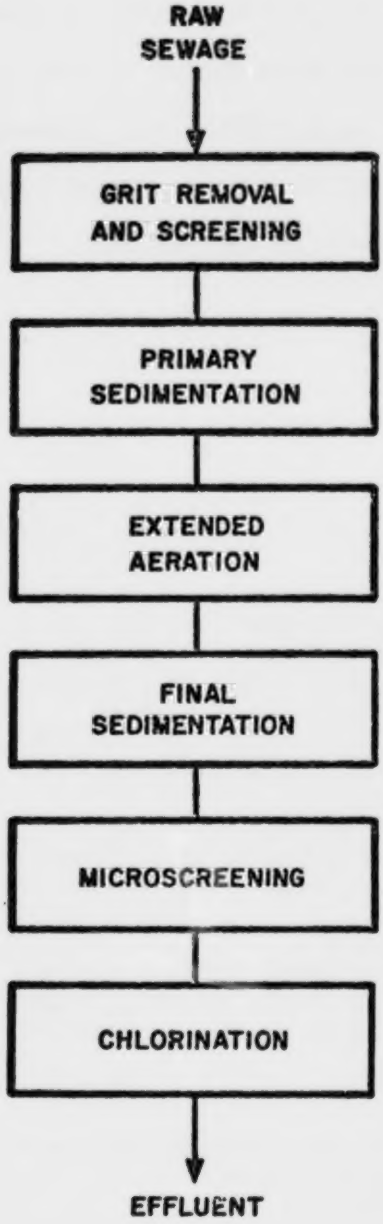
FIGURE 15

LEMONT WATER RECLAMATION PLANT SCHEMATIC DIAGRAM



F. Chlorination
   Sodium hypochlorite with chlorine contact chamber.

G. Sludge treatment
   Gravity sludge thickening tank
   Two complete mix digesters with external heat exchangers, 3200 ft^3 capacity

The aeration facility was designed to utilize the extended aeration process in order to provide nitrification as well as BOD and suspended solids removal. The extended aeration process has worked out rather well at this plant. The final clarifier effluent ordinarily averages about 5 mg/l BOD and 10 mg/l suspended solids. Effluent ammonia generally is less than 0.5 mg/l.

Table 3 summarizes the operating results for November 1972, a fairly typical month. Influent ammonia nitrogen averaged 11.1 mg/l and effluent ammonia nitrogen averaged only 0.2 mg/l, a reduction of 98 per cent. Effluent nitrate nitrogen increased accordingly by an average of 12 mg/l. Plant flow averaged 1.26 mgd, resulting in an average aeration tank detention time of 11 hours. The MLSS averaged 3460 mg/l and it's interesting that the plant has never encountered any serious settling problems. The BOD loading and F/M ratio were as would be anticipated from the extended aeration process very low.

The extended aeration process typically has long aeration times and low sludge wastage rates. Average monthly aeration times have varied from 10 hours to 26 hours. Because of low sludge production, weekly grab samples of the return sludge were analyzed for metal content to ascertain if any concentration was occurring. Table 4 shows the average monthly results from December 1972 through June 1973. There is to date no evidence of metal concentration in the sludge.

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TABLE 3

OPERATING RESULTS OF LEMONT - WRP

NOVEMBER 1972

NH_3-N Primary	11.1 mg/l
NH_3-N Final	0.2 mg/l
Percent Reduction	98%
NO_3-N and NO_2-N Primary	1.47 mg/l
NO_3-N and NO_2-N Final	13.60 mg/l
NO_3-N and NO_2-N Increase	12.13 mg/l
Flow to Aeration Tanks	1.26 mg/l
Detention Time	11 hours
BOD Loading	13 lbs/1000ft^3
MLSS	3460 mg/l
F/M	$0.05 \frac{\text{lbs BOD}}{\text{lb MLSS}}$
D.O. Final	8.2 mg/l
Temperature Final	54°F
S.S. Raw	101 mg/l
S.S. Primary	85 mg/l
S.S. Secondary	10 mg/l
BOD Raw	128 mg/l
BOD Primary	96 mg/l
BOD Secondary	5 mg/l

* Range of Detention times for November 1972 were 5 to 24 hours.

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TABLE 4

METALS IN RETURN SLUDGE OF LEMONT-WRP DECEMBER 1972 - JUNE 1973

	Zn	Cd	Cu	Cr	Fe	Ni	Pb
DEC.	1.33	0.01	0.22	0.02	5.6	0.00	0.59
JAN.	0.75	0.00	0.16	0.02	3.4	0.00	0.36
FEB.	0.53	0.00	0.14	0.01	1.7	0.01	0.27
MAR.	0.83	0.01	0.12	0.01	3.1	0.02	0.38
APR.	1.18	0.01	0.24	0.02	5.8	0.03	0.62
MAY.	0.79	0.01	0.20	0.01	4.5	0.00	0.38
JUNE	0.99	0.00	0.25	0.01	5.4	0.01	0.46
AVERAGE	0.91	0.01	0.19	0.01	4.2	0.01	0.42

All values reported as mg metal per gram return sludge suspended solids. Data represents monthly averages of weekly garb samples.

Nitrification has been maintained without interruption since June of 1972. Sewage temperatures at the plant have been as low as 45°F without any impairment in ammonia removal. Although our current interest in single-stage nitrification lies primarily in systems having shorter detention times, the data from the Lemont WRP has demonstrated that the process, when operated at long detention times, can consistently produce a well-nitrified effluent. As the flow to the plant increases due to expansion of the sewered population, we will determine what operational changes may be required to maintain effective nitrification.

Summary

As described in the preceding sections, the MSDGC past pilot plant experiences in achieving nitrification of its domestic wastewaters has been with the two-stage process. Also, the District's 50 mgd Salt Creek water reclamation plant which is currently under construction is designed as a two-stage plant. However, it should be pointed out that the MSDGC policy toward nitrification at any future plants or in the expansion of present plants is still not completely resolved. Currently, the concept of single-stage nitrification at all the major plants is being tested on a pilot scale. In particular, a pilot plant currently under construction at West-Southwest will have a maximum flow capacity of 26 mgd, (detention time, 5.5 hours) with the return sludge flow capability of 26 mgd.

Nitrification of a High Ammonia Content Sludge Supernatant by Biological Processes

As part of its land reclamation and sludge recycle program, the MSDGC has been barging anaerobically digested sludge to Fulton County, Illinois for application to strip-mined lands to promote row crop production and grazing. Before the digested sludge is spread on the land, it is held in large holding basins or lagoons. Since the sludge remains in the lagoons for a significant period of time, much compaction of the digested solids occur and a substantial layer of supernatant, which is relatively low in suspended solids, is formed. Because of the direct influence of high nitrogen content on land demand, it was imperative that the MSDGC develop methods for reducing the ammonia content of the sludge lagoon supernatant (S.L.S.), since the ammonia content of the S.L.S. constitutes up to 50% of the total nitrogen content of the lagooned digested sludge.

Basically, there were two alternatives for reducing the ammonia content of the S.L.S. Either the S.L.S. could be barged back to the West-Southwest Treatment Plant in Stickney, Illinois or the S.L.S. could be treated on site at Fulton County with the effluent being discharged to the local waterways.

The overall goal of the investigation was to pursue the second alternative and determine the feasibility of biologically treating S.L.S. for stream discharge. Because of the small amount of information available on the biological treatment of a high strength ammonia waste, particularly sludge supernatant, it was of importance to initially establish the feasibility of biologically nitrifying S.L.S. Two methods of biological treatment were evaluated: namely, a conventional activated sludge or biological slurry system and a system utilizing partially submerged rotating discs. The use of a rotating disc system was investigated because of its potentially low maintenance and operating costs.

Throughout the study, S.L.S. from Fulton County was shipped periodically by truck to the W-SW Treatment Plant where the pilot tests were conducted. As can be seen in Table 5, which lists the chemical characteristics of the S.L.S. which was used in the study, S.L.S. is a high strength ammonia waste water with the NH_3-N averaging 547 mg/l. Since the oxygen required to biologically oxidize ammonia to nitrates is 4.6 lbs O_2/lb NH_3-N oxidized, the theoretical oxygen demand is considerable and can be as high as 4050 mg/l (when 879 mg/l of NH_3-N is oxidized to nitrates). It can also be seen in Table 5 that there was on the average only about 3.3 lbs alkalinity per lb of NH_3-N contained in the S.L.S.

Since the oxidation of NH_3-N involves the theoretical consumption of 7.2 lbs of alkalinity per lb of NH_3-N oxidized, the S.L.S. did not contain enough alkalinity to meet the demand. This was important because the autotrophic bacteria which oxidize ammonia require a pH of at least 6.5 - 7.0 in order to

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TABLE 5

CHEMICAL CHARACTERISTICS OF SLUDGE LAGOON SUPERNATANT FROM FULTON COUNTY

CHEMICAL PARAMETER	AVERAGE	RANGE
pH		8.0-8.4
Alkalinity, mg/l	1834	1215-2843
Total Kjeldahl Nitrogen (TKN), mg/l	703	465-975
NH_3-N, mg/l	547	292-879
NO_2-N, mg/l	0.21	0.04-0.86
NO_3-N, mg/l	0.40	0.26-0.74
Total Solids,mg/l	1190	813-1402
Total Suspended Solids, mg/l	150	52-231
Volatile Suspended Solids, mg/l	97	39-126
Total BOD	64	28-121
Total COD	753	345-1119
Total PO_4-P	40.5	23.1-51.1

function properly. If the pH goes below 6.5, the rate of oxidation drops off radically. Therefore, throughout the study, the pH was controlled in the range of 7.8 - 8.2 with the addition of sodium carbonate.

The rotating disc pilot system consisted of a semi-circular fiberglass tank which was divided into four equal volume compartments. The tank was equipped with 2' diameter plastic discs arranged on a horizontally rotating shaft. The total surface area provided by the discs was 250 ft^2. The volume of the unit was 35 gallons, with the discs submerged 40% of the diameter. Disc rotational speed was maintained at 3 rpm for the duration of the study.

The slurry system consisted of a 50-gallon plexi-glas aeration tank, which was separated into seven compartments to simulate a plug-flow system, and a 6' high clarifier which provided a maximum surface area of 2.25 ft^2. Air was supplied by diffused aeration through porous membranes.

Before the continuous flow operation began, both units were initially filled with S.L.S. diluted by 50% with tap water. The bio-mass within each system was allowed approximately two weeks of acclimation before continuous flow operation began. The temperature throughout the study in both units was maintained in the range of 22^o - 28^oC.

As can be seen in Figures 16 and 17, which show the performance of both the rotating disc system and the slurry

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FIGURE 16

PERFORMANCE OF THE SLURRY SYSTEM

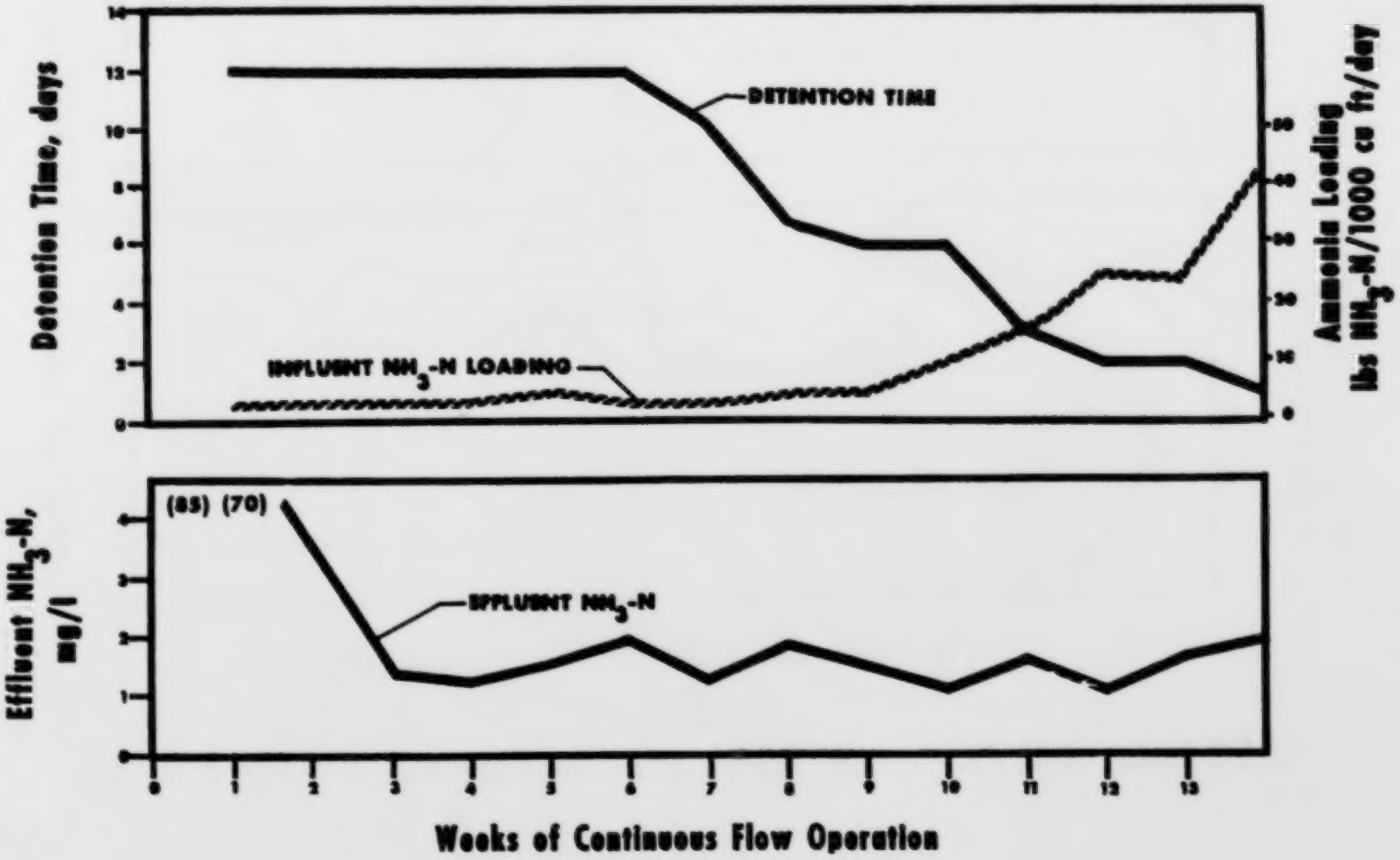

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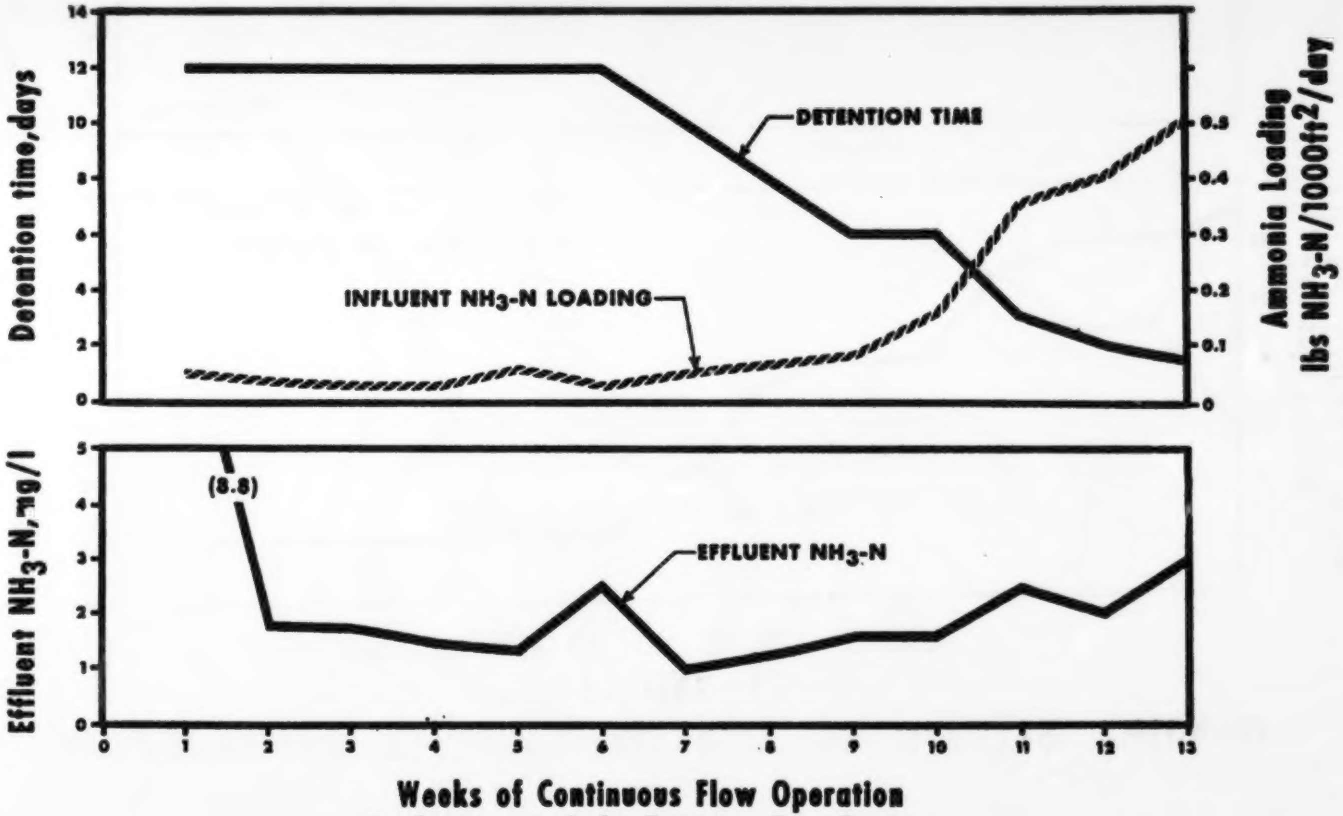
FIGURE 17

PERFORMANCE OF THE ROTATING DISC SYSTEM



Performance of the Rotating Disc System

system, respectively, the S.L.S. was highly amenable to biological nitrification. Except for the first two weeks of operation, both systems performed extremely well through the progressively higher ammonia loadings applied. The average effluent NH_3-N at the highest loadings for the rotating disc system and slurry system was 2.5 mg/l and 2.0 mg/l, respectively. Except for the first few weeks when some ammonia was stripped in both units (as determined by a nitrogen balance), the effluent NO_3-N and NO_2-N generally exceeded the ammonia removed, which indicated a high amount of nitrification and conversion of organic nitrogen to NH_3-N and subsequent oxidation to NO_2-N and NO_3-N. The overall total Kjeldahl nitrogen removal (TKN) for both systems at the highest loading was approximately 99%. Early in the study, there was some inhibition of the NO_2-N oxidation in both systems as the effluent NO_2-N generally was much greater than the effluent NO_3-N. However, at detention times of ten days and lower, the effluent NO_2-N was generally less than 1.0 mg/l.

The air flow rate to the slurry system was controlled to maintain a D.O. of at least 1 - 2 mg/l in each of the compartments. The D.O. in the rotating disc system was at least 2.0 mg/l in each compartment, even at the highest ammonia loadings. The amount of Na_2CO_3 required to maintain a pH of 7.8 to 8.2 in the rotating disc system averaged about 3 - 4 lbs of alkalinity per lb of NH_4-N oxidized, while the slurry system averaged

4 -5 lbs of alkalinity per lb of NH_3-N oxidized.

In the temperature range of the study (22-28°). it can be concluded that high rates of ammonia oxidation in S.L.S. can be achieved by both the slurry and rotating disc system as long as the pH is properly maintained with Na_2CO_3. Further work is being conducted to determine the effects of lower temperatures on both processes. Also, a better assessment of the effluent quality in terms of BOD and suspended solids of both systems at the colder temperatures will likewise be investigated.

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STORM AND COMBINED SEWER
ABATEMENT TECHNOLOGY
IN THE UNITED STATES
- AN OVERVIEW -

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I. Introduction

The Office of Research and Development of the U. S. Environmental Protection Agency is organized so that program responsibility to develop new technology for treatment and control of water pollution caused by urban runoff falls within two organizational units. The Municipal Pollution Control Division of the Washington Headquarters staff is responsible for planning, coordinating and assessing the program. The Advanced Waste Treatment Laboratory in Cincinnati, Ohio, is responsible for implementation. There are, of course, other activities within these units but this seminar will be limited to urban runoff pollution abatement technology.

The following are the principal sources of runoff induced pollution as viewed in our program activities: combined sewer (domestic sewage and runoff) overflows, sewered storm water discharges, sanitary sewers surcharged by infiltration, treatment works or pump station bypass and urban non-point or overland runoff.

Given these sources it follows that the specific causes are precipitation and snow melt resulting in hydraulic overload relief of combined wastewater collection and transport networks, pollutants flushed from urban surfaces and bypass of excess wastewater flows at treatment facilities to prevent process upset.

When the causes are considered in this context, it is evident that remedial solutions must include elements of urban hydrology as well as sanitary engineering. The superimposing of urban hydrology on the usual sanitary engineering solution methods broadens the scope of problem definition and abatement approach. Considering the situation in such a manner enters into the realm of metropolitan water balance systems or the total system concept.

Consideration of the total system in problem definition and alternate solution schemes is somewhat new to the practicing sanitary engineer in the United States who traditionally views waste water treatment as the principal means of water pollution abatement. Treatment of foul water will always be necessary but improved technology development and application in prevention, reduction and control of hydraulic and pollutant factors must be given high priority when addressing storm and combined sewer overflow pollution sources.

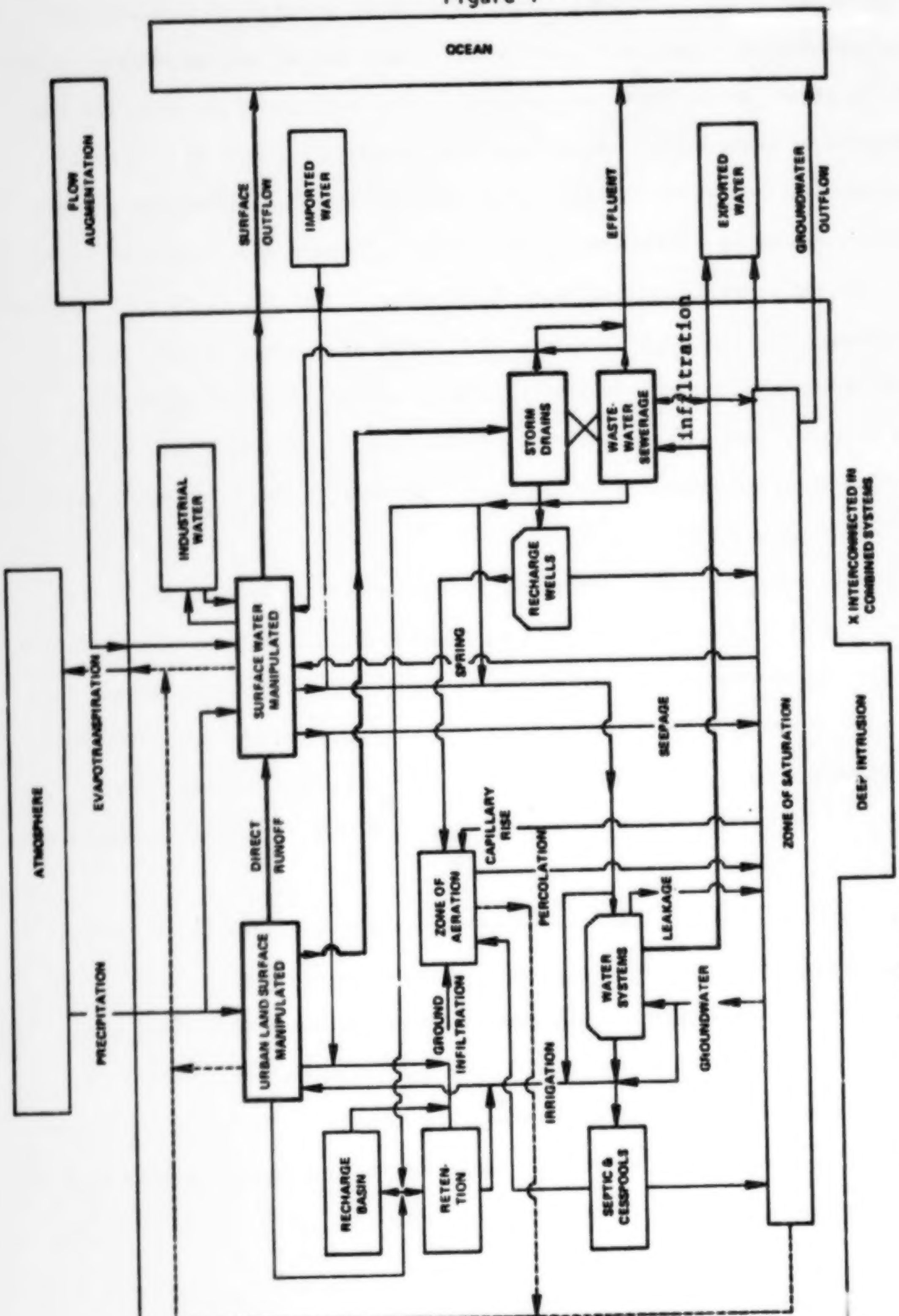
Figure 1 is a schematic display of the total metropolitan hydrologic system which must be examined in developing data and applying technology for urban water quality improvement. It is illustrative of the scope in cause and effected uses which should be considered in metropolitan runoff pollution.

Figure 1 may be used for outlining areas for which water quality and quantity data must be obtained to assess relative significance or as a guide to where applied research will have the most effect.

For example, the conditions of the urban land surface effects the quality and quantity of: direct runoff (non-point), storm drainage discharge and, where interconnected in combined systems, the normal domestic sewage flows. Therefore, as a first cut in prevention one would evaluate the possibilities and effectiveness of manipulating urban land surfaces in reducing water pollution.

The Figure also serves as an example of the need to apply the system concept in the analysis, planning and implementation of a Research, Development, Demonstration Program.

Figure 1



URBAN HYDROLOGIC SYSTEM
(ADAPTED FROM: "SUMMARY OF THE HYDROLOGICAL SITUATION IN LONG ISLAND, N.Y., AS A GUIDE TO WATER MANAGEMENT ALTERNATIVES", BY O.L. FRANKE AND N.E. McCLYMONDS, U.S. GEOLOGICAL SURVEY PROFESSIONAL PAPER 627F, 1972)

This last point is of importance. The large majority of research and development efforts and resources in the United States, and apparently other Nations of the world, have been devoted to the development of tools for the treatment of wastewaters. Relatively little effort has gone into overall problem solutions which include prevention and control. Within our program attempts are being made to change the thrust from tool development to storm and combined sewer problem solution development. That is, we hope to complete preliminary National field measurement programs that when coupled with performance data from categories of tools, i.e., liquid-solid separation, flow control, rate attenuation, etc., will provide design criteria as dictated by requirements of performance or by desired receiving water quality.

II. Problem Description

In the paper entitled, "Municipal Pollution Control Technology in the United States of America" presented in Tokyo by Mr. Frank M. Middleton during the 1971 Conference, it was noted that the single dissimilarity between the municipal wastewater treatment programs of Japan and the United States was the trend in Japan of continuing to build combined sewer systems. In comparison, current construction in the United States is primarily separate systems.

There is evidence of some change beginning to take place in the United States away from the trend of constructing separate systems.

In cities where existing collection networks have portions which are combined, consideration is now being given to continuing constructing combined sewers as the collection network is expanded. The pollution loads generated in urban runoff and the possible future need to reduce this source is the motivating force for the change in trend.

An example is the recommendations from the recent comprehensive report of the Department of Public Works of San Francisco. This exhaustive investigation determined that to attain the desired water quality in San Francisco Bay, it would be necessary to treat a large portion of the urban storm water runoff. Therefore, one of the principal recommendations of the report is to continue to construct combined sewers predicted on the control and treatment of wet weather flows.

To gain a perspective of the problem in the United States a brief summary of the physical and pollutional aspects of the storm and combined sewer discharges will be given. Tables 1 and 2 give the magnitude with respect to population distribution and length of sewers.

In the combined interceptor sewers the ratio of wet-weather to dry-weather capacity ranges from 1:1 to 8:1. The median ratio is 4:1. This, however, is somewhat misleading in that the rain events over the urban areas are usually of the cell patterns and short, intense precipitation in a catchment area will cause overflow at one interception point while unused capacity is available in other portions of the system.

Several of our projects have documented that rainfall intensities as low as 0.01 in/hr (0.025 cm/hr) cause overflow from combined sewers. Further, common rainfall intensities of 1.0 in/hr (2.5 cm/hr) will cause flow rates of 50 to 100 times the dry-weather flow at the interception point.

The incidence of overflow events for each relief point is about 30 per year. Consideration of an entire metropolitan area, however, yields a range of 60 to 100 events per year for the entire combined collection network. The average duration of overflow is about 5 hours. In storm sewer discharges the average number of events per year is 95 with an average duration of 7 hours.

Table 1

Number, Population and Area of U.S. Communities Served by Combined Sewers

Population Groups	Total U.S.	Number Served by Cmb. Sewers	Population (in 1,000's) Served by Cmb. Sewers	Area (in 1,000's) Served by Cmb. Sewers		
				Acres		Ha
over 10^4	132	75	26,261	1,423	-	576
50,001 - 100,000	201	86	3,854	472	-	191
25,001 - 50,000	432	119	2,385	313	-	126.7
10,001 - 25,000	1,134	203	1,865	337	-	136.5
5,001 - 10,000	1,394	227	911	214	-	867
1,001 - 5,000	6,613	458	874	235	-	955
1,00 and less	9,874	161	86	35	-	142
	19,780	1,329	36,236	3,029	-	1

Of the total sewered population in the United States approximately 29 percent is served by combined sewers.

The breakdown in round figures is: unsewered 74 million, separate sanitary 90 million, combined sewers 36 million.

Table 2

Separate Sanitary		Combined Sewers		Separate Storm	
Miles	Km	Miles	Km	Miles	Km
53,801	86,566	56,132	90,316	21,571	34,708

(Network = 120 events)

There are 14,200 plus combined sewer relief points and an estimated 700 storm sewer discharge points. In the combined portion 72% have an interval diameter of 24 inches (61 cm) or less. There are 10,000 plus regulators in the combined networks.

Another factor to be considered in examing the physical characteristics of combined sewer systems is that there are many cross overs between separate sanitary and combined networks which are unrecorded and that many storm systems are laid at shallow depths above the separate sanitary lines with frequent exfiltration from the storm sewers and infiltration into the separate sanitary sewers. Such infiltration and illegal inflow connections result in many miles of sewers which are identified as separate to be, in fact, acting as combined sewers.

When determining the receiving water pollution loading from combined sewer overflows, several considerations not normally applied to dry-weather domestic flows must be taken into account.

A summary is as follows:

A. Both strength and total mass emission of pollutants on an event basis be considered in determining receiving water degradation potential.

B. The characteristics and concentrations of the polluting parameters change with rate (cfs-l/sec), time from start of discharge and real time of occurrence.

C. The rate of polluting parameter strength decrease is less than the rate of hydraulic or flow increase, therefore, the total emission of pollutants increases as the overflow intensity increases.

D. There is, under the more common conditions, little dilution of the slug load in the receiving water during the initial critical time period. This includes flowing streams as well as lakes, bays and estuaries.

E. In flowing streams dissolved oxygen deficits are usually more severe at greater distances downstream of the overflow point source than the distance and severity during normal treatment works operation. This is caused by the slug load reaction time, the decrease in reaeration rate due to increase volume in the stream, and increased flow rate.

F. The oxygen demanding characteristics of combined sewage cannot be quantified in the same manner as dry weather sewage. For example, the heavy metals from the street flush portion inhibit the microbiological action in the BOD analysis. Therefore, the BOD_5 reading is not a valid indicator of oxygen demand potential in combined sewer overflows. Additionally, oxidation of the heavy metals, which can occur at a high rate, is masked in the standard BOD analysis.

G. There is greater differences in the vertical distribution of pollutants in overflow relief flows than in normal dry-weather flow. Sampling procedures are more critical in accurate determination of pollution loads.

These observations apply to storm water discharges as well.

Tables 3 and 4 indicating pollutant concentration and loads for combined and storm sewage discharges are given for comparative purposes.

Table 3

Comparison of Quality (Strength) of Combined Wastewater Overflows

Type of Wastewater and City	BOD_5 mg/1 Ave.	COD mg/1 Ave.	SS mg/1 Ave.	Total Coliform MPN/100 ml.
Raw Sanitary Flow	200	500	200	$5x10^7$
Primary Effluent	135	330	80	$2x10^7$
Secondary Effluent	25	55	15	$1x10^3$
Combined Sewer Overflows				
Atlanta, Ga.	100	-	-	$1x10^7$
Berkeley, Calif.	60	200	100	-
Brooklyn, N.Y.	180	-	1,051	-
Bucyrus, Ohio	120	400	470	$1x10^7$
Cincinnati, Ohio	200	250	1,100	-
Des Moines, Iowa	115	-	295	-
Detroit, Michigan	153	115	274	-
Kenosha, Wisconsin	129	464	458	$2x10^6$
Milwaukee, Wisconsin	55	177	244	-
Racine, Wisconsin	119	-	434	-
Sacramento, Calif.	165	238	125	$5x10^6$
San Francisco, Calif.	49	155	68	$3x10^6$
Washington, D.C.	71	382	622	$3x10^6$

Table 4

Comparison of Quality (Strength) of Storm Water Discharges

City	BOD_5 Ave.	COD Ave.	SS Ave.	Total MPN/100 ml
Ann Arbor, Michigan	28	-	2,080	-
Des Moines, Iowa	36	-	505	-
Durham, N.C.	32	224	-	3x105
Los Angeles, Calif.	9.4	-	1,013	-
Madison, Wisconsin	-	-	81	-
New Orleans, La.	12	-	26	$1x10^6$
Sacramento, Calif.	106	58	71	8.10^5
Tulsa, Oklahoma	11	85	247	$1x10^5$
Washington, D.C.	19	335	1,697	$6x10^5$

For comparison purposes the following generalizations with respect to pollution loadings can be made although wide variations can exist at specific locations.

A. For any given collection system the raw domestic sewage which is carried out of the combined sewers by excess flows is about 5 to 7 percent of the total dry weather flow on an annual basis. With respect to time, the accumulated overflow events total about 4% or 14 1/2 days per year.

B. On an annual basis the pounds of BOD_5 discharged in combined sewer overflows is approximately equal to the pounds discharged from a secondary treatment plant serving the combined collection area.

C. On an event basis the organic loads placed on the receiving water during combined sewer overflow are 8 to 12 times the secondary treatment plant effluent load.

D. Urban runoff is similar to dilute industrial wastewater in many respects. Calculations indicate that a moderate to heavy storm over an urban area will wash more oxygen demanding load into the receiving water than the same area's raw sanitary sewage during the same period of time.

The impact of storm generated pollution on receiving waters is difficult to quantify with precision. The relative significance of various runoff pollution sources in a reach of water requires a long period of sampling under a sufficient range of the primary variables.

However, it does appear the myth of dilution can be put to rest. The high frequency, low intensity storms are the greatest offenders. The runoff volume is low so that with respect to the receiving water volume there is little percent change. But the pollutant concentrations are so high that significant levels of change are noted in the receiving water.

For example, rainfall intensities up to 0.4 in. (10 cm.) per hour account for 98 of the average 120 runoff events per year but only 32 percent of the total volume of runoff. Yet these low intensity events account for over 85% of the pollutant mass emission.

Therefore, if there is a dilution factor it occurs only in the less frequent long duration, constant precipitation rate storms.

This is a brief outline of the extent, cause, source and incidence of the stormwater runoff problem. The one important and unanswered question is: What ranking value iwth respect to cost effectiveness will we give the problem relative to other sources of water pollution, e.g., industrial or municipal treatment works? This of course would assign the priority of remedial construction funding. Attempts to resolve this question are currently underway.

III. Strategy for Research and Development

The Federal Water Pollution Control Act Amendments of 1972 contains sections which deal with technology research and development, urban runoff, consideration in developing area wide plans, and permit conditions for combined sewer overflows. There are requirements for quantifying the pollution of stormwater runoff. There is, however, a large gap in current construction funds for remedial hardware. The relative significance of the problem must be determined with a high degree of certainty before immediate remedial funding allocations can be justified.

Future corrective action, however, cannot be denied. As the conventional pollution sources of municipal treatment works and industrial plants are brought under control, the significance of combined and stormwater discharges will increase. In many cases, the receiving water quality standards will not be attained until these sources are reduced or eliminated.

With the strong possibility of near future construction expenditures, it would be well to review the strategy and tools developed for abatement.

The formulation of the program strategy for research and development activities was influenced by four elements. These can be identified from Figure 2. They are: (1) the requirements of the water pollution control act, (2) the existing conditions, (3) the level of technology practiced along with the State-of-the-Art of the technology, and (4) the funding available.

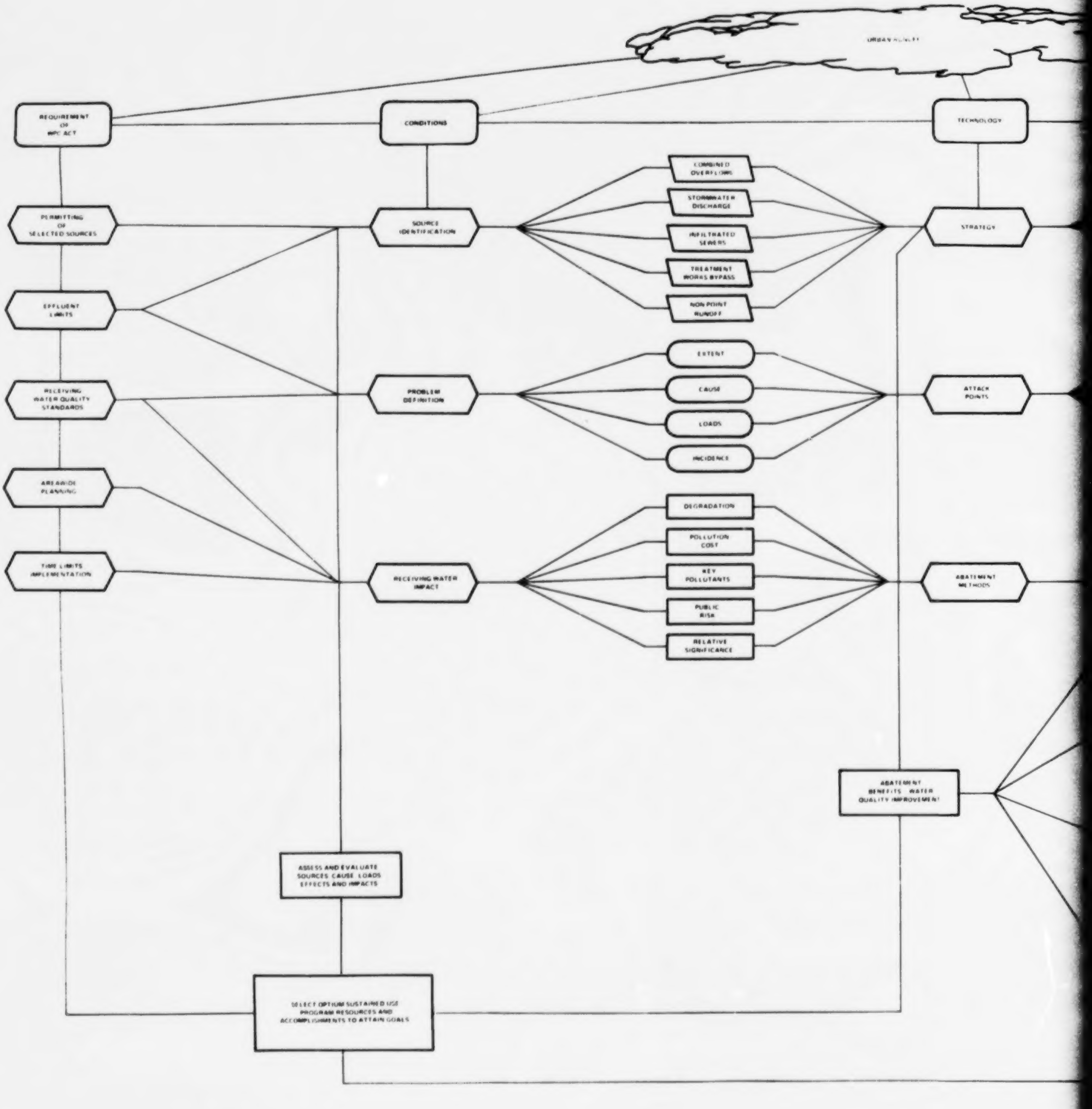
Initially, the strategy of the program, which began eight years ago, was to carry out full-scale demonstration projects of new and improved abatement tools wherein design, performance, benefits and cost data would be developed. The goals of this program have been partially realized. Over the first few years of the program emphasis was changed to other areas of environmental pollution and fewer full-scale projects could be supported than had been originally planned. As feedback from demonstrations was obtained, the thrust of the program changed, as mentioned in the Introduction, to methods of problem solution.

However, there have been several pilot and bench top projects completed, along with some full-scale demonstrations in development of novel or improved tools.

IV. Methods of Abatement

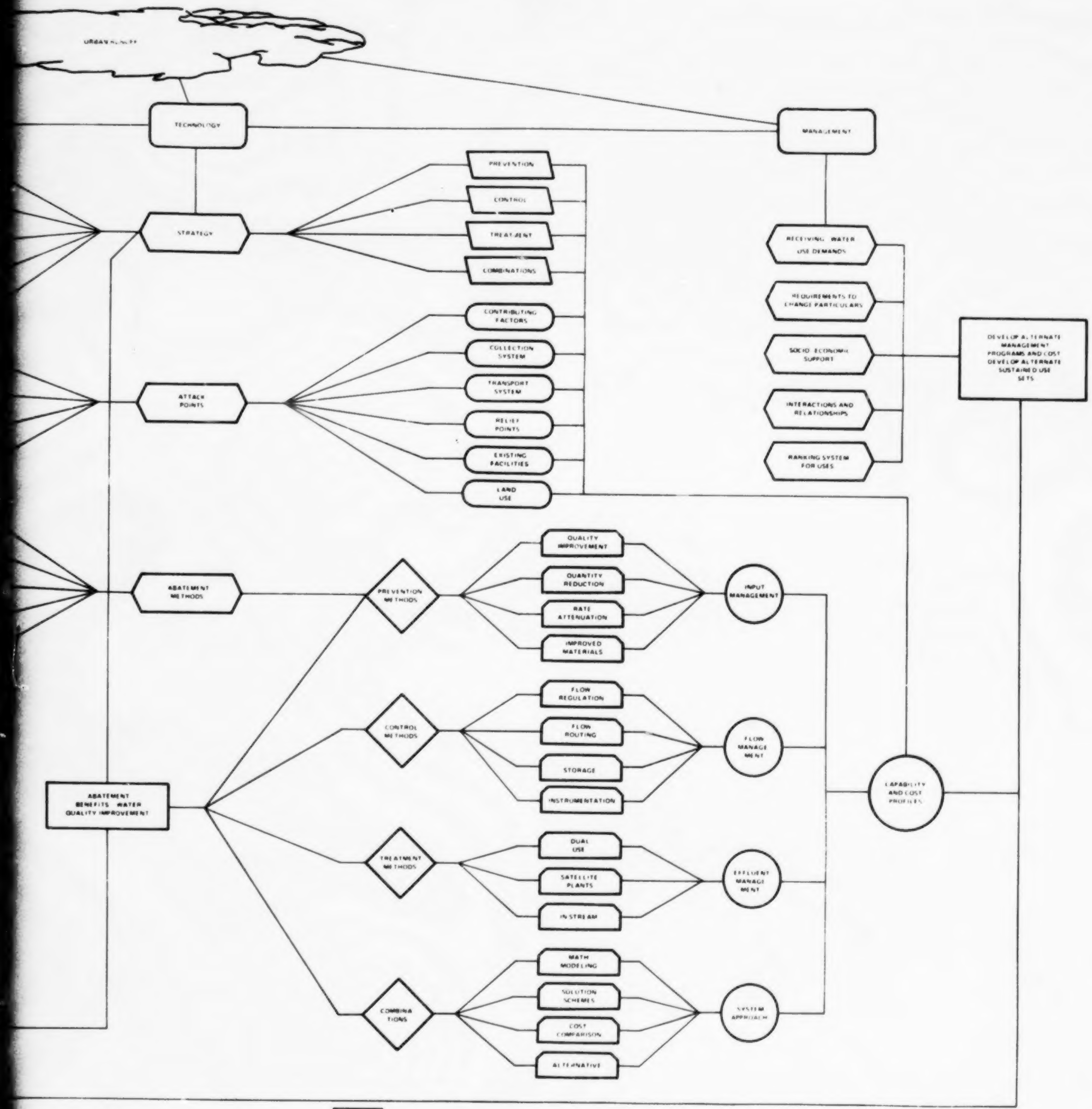
A brief summary of selected projects grouped by principal methods as shown in Figure 2 will be given. It should be noted that many individual projects cut across the category divisions and may incorporate two or more

Figure 2


URBAN RUNOFF
REQUIREMENT OF WPC ACT
CONDITIONS
TECHNOLOGY
PERMITTING OF SELECTED SOURCES
EFFLUENT LIMITS
RECEIVING WATER QUALITY STANDARDS
AREAWIDE PLANNING
TIME LIMITS IMPLEMENTATION
SOURCE IDENTIFICATION
PROBLEM DEFINITION
RECEIVING WATER IMPACT
COMBINED OVERFLOWS
STORMWATER DISCHARGE
INFILTRATED SEWERS
TREATMENT WORKS BYPASS
NON POINT RUNOFF
EXTENT
CAUSE
LOADS
INCIDENCE
DEGRADATION
POLLUTION COST
KEY POLLUTANTS
PUBLIC RISK
RELATIVE SIGNIFICANCE
STRATEGY
ATTACK POINTS
ABATEMENT METHODS
ABATEMENT BENEFITS WATER QUALITY IMPROVEMENT
ASSESS AND EVALUATE SOURCES CAUSE LOADS EFFECTS AND IMPACTS
SELECT OPTIUM SUSTAINED USE PROGRAM RESOURCES AND ACCOMPLISHMENTS TO ATTAIN GOALS

A

Figure 2



Storm and Combined Sewer
Pollution Control Program

B

methods. However, for purpose of presentation, every project has been assigned to one remedial method category.

A. Prevention or Input Management

Fourteen projects have been completed in the category. The projects include: (1) quantity reduction (infiltration control), (2) quality improvement (deicing methods), (3) runoff rate attenuation (porous pavements), and (4) improved materials (electromagnetic sub-surface profiling).

The rational of the objectives in this remedial method grouping is best illustrated by referring to Figure 1. As mentioned, urban land surface conditions effect the quality and quantity of non-point runoff, storm discharges and combined sewer overflows. And, as importantly, they effect the time of concentration. Therefore, any manipulation or improvement of the urban surface conditions which would reduce the volume of runoff or increase time of concentration or improve the runoff quality is considered a prevention method.

An example of work in this area is the feasibility study entitled "Investigation of Porous Pavements for Urban Runoff Control". Various methods of construction to obtain permeable surface covering were examined in the investigation. The material which indicated the highest possibility of success is porous asphaltic concrete roadways.

The reason for selecting permeable street surfaces is that the urban street or roadway acts as a collector, a transport pathway and generator of pollutants. Its use, physical characteristics, location and method of construction makes a transition or hinge point between air, land and water pollution. Therefore, a likely pollution control attack area is improvement in design, method of construction and operation.

Viewed in this fashion there are many benefits. For example, if a permeable surface is successful, part of the runoff will enter the ground, reducing the volume and recharging urban ground waters. Part will flow through the base material and thus be rough filtered and the time of concentration will be increased. Therefore, we may realize both prevention and control at a critical point in the hydrologic system.

Based on the feasibility investigation there are two current demonstration projects, recently initiated, in which design and performance data are to be gathered on porous asphaltic parking lots and low design residential streets. Other projects are being considered in demonstration of improved curb and gutter design to facilitate street cleaning and swirl separators at inlets, in place of catch basins, to remove particulates (the majority of surface runoff pollutants) by concentrating them in a relatively small percentage of the runoff volume followed with storage and treatment.

B. Control or Flow Management

Ten projects have been completed in this method of abatement. This is a somewhat misleading indication of the effort which has gone into the category. Many of the effluent management or treatment projects have significant work in control of flows. The projects listed here are totally dedicated to flow management. The projects include: (1) flow regulation (a manual of practice), (2) flow routing (a dispatching system), (3) storage (deep tunnels), and (4) instrumentation (maximizing in-system capacity).

A capability profile will be given on the deep tunnel concept, therefore, an example in this category will not be discussed. However, a few words on the overall rational are in order.

Flow control in combined sewer pollution abatement appears to be the most cost effective approach providing, of course, there is adequate dry weather treatment facilities. In the United States very few communities with combined sewers, in whole or in part, know how their collection-transport systems react to runoff events. In all of our problem definition and demonstration projects it has been found that either through flow routing, improved regulation or in-system storage, the rate of flow at a given relief point could be controlled to some degree thereby reducing the volume and incidence of raw sewage discharge. As an added benefit, the efficiency of the treatment works can be markedly improved.

Our conclusion is that a well designed sewer monitoring and sampling program integrated with in-depth rainfall and runoff monitoring to obtain accurate basic data for use in the Storm Water Management Model will return many times over the original investment when an abatement scheme is to be chosen and construction started.

C. Treatment Methods or Effluent Management

The first groupings in this category are: (1) dual or multi use facilities, (2) satellite plants or treatment at outfall points, and (3) in-stream treatment.

Each of these groupings could be broken down into similar specific processes such as: physical, physio-chemical, bio-physical, bio-chemical and chemical.

The total of thirteen projects have been completed in this category. A brief summary of successful treatment modes developed to pilot stage is as follows:

1. Physical:
    - (a) sedimentation and flow control (combined sewers)
    - (b) screening, dissolved air flotation (combined sewers)
    - (c) deep bed, dual media filtration (storm runoff and combined sewers)
    - (d) swirl separation (combined sewers and storm runoff)
    - (e) microstraining and disinfection (combined sewers)
2. Physio-Chemical:
    - (a) powdered activated carbon (combined sewers)
    - (b) coagulation - flocculation aids (combined sewers)

      Item (b) is in conjunction with physical mode projects in (1) above.
3. Bio-Physical:
    - (a) detention with bio action and physical separation, includes dual use (combined sewers)
4. Biological:
    - (a) dual use contact stabilization (combined sewers)
    - (b) dual use trickling filter (combined sewers)

5. & 6. Bio-Chemical and Chemical Modes have not been evaluated.

Although bio oxidation through storage lagoon aeration are among on-going projects, the operating data are not available. In the 5 and 6 Subgroupings, the completed work is principally in methods of disinfectant generation which have been modified to suit combined sewer overflow treatment. Two methods for generating sodium hypochlorite and one for chlorine dioxide have been developed.

Abstracts of completed projects are in the Appendix. A capability profile will be given on a dual use, biological demonstration facility.

As a final observation in this category, it appears that dual use, in a physio-chemical or bio-physical mode of operation, is the most cost effective treatment scheme and that in-stream treatment is a poor risk for effectiveness.

D. Combinations or System Approach

In this category, the prevention, control and treatment methods are brought together and mathematical modeling is added to develop solution schemes, make cost effectiveness comparisons and examine alternate solution methods. It is here where the total system approach is developed. In all, six projects have been completed in this category.

Our primary vehicle in approaching the total system concept is the Storm Water Management Model. A capability profile will be given on this project.

The remaining completed projects in this category are principally engineering investigations of urban areas which develop methods of examining alternate schemes and cost effectiveness comparisons. From these, we hope to arrive at a uniform means of: (1) predicting pollution loads, (2) gathering and presenting evaluation data, and (3) arriving at the true cost and receiving water benefits.

V. Capability Profile: Effluent Management

A. Dual Use Biological Treatment

The concept of this project is to modify an existing activated sludge process in such a manner so as to provide contact stabilization for excess wet weather flows which are being bypassed at the plant to prevent process upset. The following discussion and tables are extracted from a paper by Charles A.Hansen and Robert W. Agnew of Rexnord Company on an EPA demonstration project sponsored by the Municipal Pollution Control Division.

Before construction of the demonstration system, the treatment train was grit removal, primary sedimentation, aeration (activated sludge), sludge mixing for anaerobic digestion, sludge thickening, clarification of the mixed liquor and chlorination. The original plant treatment capacity was 23 MGD (1007 l/sec-). The interceptor size at the plant wet well is 72 in. (183 cm.) with a 50 MGD (2190 l/sec) capacity. Bypasses of excess flows were frequent.

A maximum design flow of 20 MGD (876 l/sec.) was selected for the demonstration project. The total wet weather capacity then being 43 MGD (1890 l/sec.).

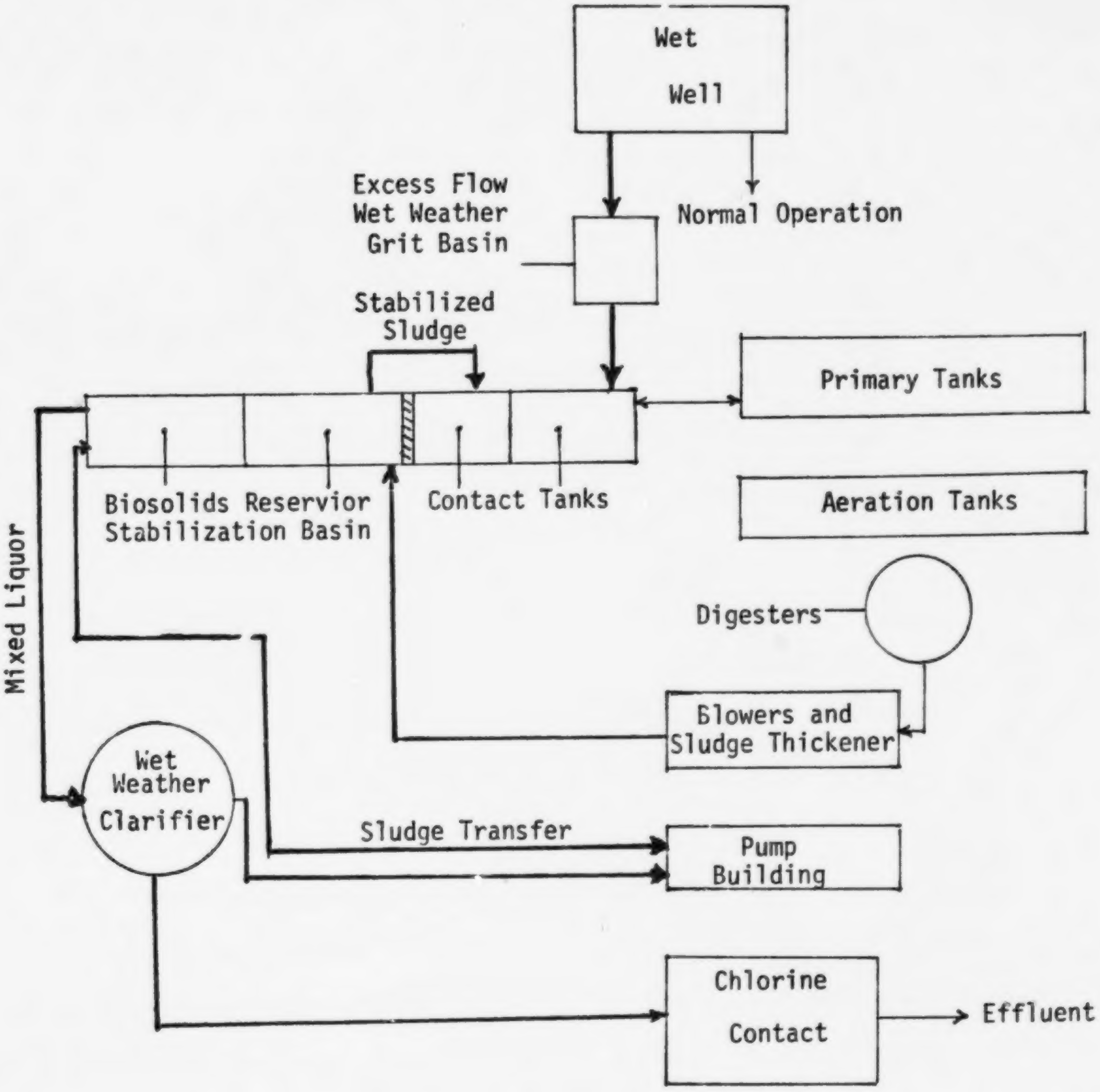
The demonstration system was designed for integration with the original facilities. A constant portion of activated sludge from the dry-weather operation is stored by balancing the biosolids reservoir inflow and outflow. The stored material is used as a source of

Figure 3

Kenosha, Wisconsin

Contact Stabilization Demonstration

Flow Diagram



biological solids in the contact stabilization treatment of the excess flows. This mode of treatment is also identified as physical adsorption, biosorption and sludge reaeration. All refer to a modification of the activated sludge process.

Modification and additions to the existing plant hardware are as follows:

1. Installation of a 20 MGD (876 l/sec.) gas driven pump in the existing wet well and diversion chamber. The pump delivers the wet weather excess flow to the demonstration system and can be used as standby for dry weather operation.
2. An unused mixing basin was converted to a grit tank. The tank is 56.5 ft. (17.23 m) long, 225 ft. (68.62 m) wide and has a mean depth of 9 ft. (2.74 m). At a flow of 20 MGD (876 l/sec.) the horizontal velocity is less than 0.2 fps (0.06 m/sec.).
3. Construction of a contact and stabilization tank in one structure divided by concrete walls into four smaller tanks. The contact section was designed to handle a maximum raw flow of 20 MGD (876 l/sec.) and a stabilized sludge flow of 5 MGD (219 l/sec.) for a 15 minute contact period. The contact section can be divided into two smaller units of different volumes to allow experiments in varying the contact time while the total flow remains constant. This also provides like raw flow pollutant characteristics while comparing contact time effectiveness. One unit of the contact tank has a volume of 164,000 gal. (620,740 l), the second 80,500 gal. (304,792 l). Therefore, by using the units either separately or simultaneously three effective volumes can

be made available. The tank has a sidewall depth of 17.5 ft. (5.34 m). Aeration is through a fixed air dispenser system along the bottom of one wall in each unit. The system can deliver up to 3,750 scfm (106.12 cu m/min.).

4. Following the contact section of the tank and divided by a concrete wall is the stabilization basin (biosolids reservoir section). This section is also divided into two parts so that various stabilization times may be evaluated. Both units of this section are 30 ft. (9.15 m) wide, 96 ft. (29.28 m) long and 17.5 ft. (5.34 m) deep. Aeration is by eight 50 hp (37.3 kw) floating mechanical surface aerators. The arrangement allows for stabilization times of up to seven days.
5. Construction of a new clarifier for use in both dry and wet weather flows. The design surface overflow rate (SOR) is 1300 gpd/sq. ft. (0.047 lps/m^2) resulted in a 140 ft. (42.7 m) diameter, peripheral feed effluent clarifier. The surface area is 15,400 sq. ft. (1420.6 sq. m) and a volume of 1.4 million gal. (5.23 million l). During wet weather operation the clarifier is isolated from the dry weather plant.
6. The necessary piping, pumps, valves and instrumentation are included.

The system is kept in readiness by maintaining a biosolids reservoir fed by the dry weather plant operation.

When excess flows occur the raw flow pump automatically starts and delivers the flows via the grit chamber to the contact section where stabilized sludge from the biosolids reservoir is proportioned in. At the same time the pump for transferring sludge from the new clarifier to the stabilization section is put into operation. The air blower for the contact tank is also started at this time. Once the demonstration system is operating the sludge detention time in the stabilization tank is short. The time being dependent on the flow rate entering the dry weather plant and the rate of transfer to the contact section. The sludge flow rate to the contact section is a function of the rate of raw combined sewage flow through the contact tank. Meanwhile, at the new clarifier the flow from the dry weather plant is diverted to the original clarifiers and the return sludge line to the dry weather plant is shut, the return sludge line from the new clarifier to the stabilization tank is opened. The clarifier then receives the feed from the contact-stabilization tank. The necessary synchronism of blower and chlorine feed increase is manually started.

When the high flow condition has subsided the wet weather system is manually taken out of service at the main control board. After shutting down the wet weather mode of operation, the start-up procedures are automatically reversed and the plant returns to normal dry weather operation. The grit and contact tanks are then emptied and the system made ready for the next event.

Table 5

Summary of Operating Conditions

Variable	Unit of Measure	Range	Mean	Metric
Gal treated	mil gal	1.789-7.558	4.007	$15.17X10^6$ L
Average flow rate	mgd	13.3-19.4	16.6	727.2 L/sec.
Duration of run	hours	2.7-11.0	5.8	
Return sludge	% of raw flow	10-55	37	
MLSS	mg/1	975-5370	3060	
Contact time	minutes	12.1-19.6	15.5	
Reaeration time	hours	1.0-7.3	2.7	
Stabilization time	days	0.5-15	4.4	1.99 Kg
F/M	1b BOD/day 1b MLSS	0.64-5.3	2.7	
Stabilization tank turnovers		0.53-6.8	2.9	
Clarifier SOR	gpd/ft^2	864-1257	1078	305 cu m/min/ha
Clarifier detention time	hours	1.3-1.85	1.46	
Clarifier turnovers		1.7-8.2	3.9	

Table 6

1972 Removal Summaries 23 Events, 92 Mil Gal. Treated

	Arithmetic Mean		Weighted Mean	
	Raw[1]	Final[2]	Raw[1]	Final[2]
Total Solids	706	471	704	455
T.V.S.	270	143	270	140
Suspended solids	325	24.7	314	26.4
S.V.S.	126	13.9	121	15.2
Total BOD	115	18.6	102	17.8
Dissolved BOD	27.6	8.5	24.1	7.6
Total organic carbon	121	24.0	113	22.8
Dissolved organic carbon	25.4	16.9	21.8	15.3
Total Kjeldahl as N	12.9	6.7	11.0	5.5
Total PO_4 as P	5.1	2.7	4.8	2.4

	Geometric Mean	
	Raw[1]	Final[2]
Total coliform count/ml	34,786	2,8[illegible]
Fecal coliform count/ml	2,308	374

[1] Raw flow sample taken after grit removal.

[2] Final sample taken prior to chlorination.

The total construction cost for the system was \$1.1 million. The capital cost breaks down to \$917 per acre (.405 ha) served or \$55,000 per MGD (43.808 l/sec.). Operating and maintenance cost cannot be determined because all of the runs have been of experimental nature and no standard procedure has been set. Future plants can be constructed at lower cost because provisions will not be necessary for operating at different levels for each variable.

The principal conclusions are:

1. The use of contact stabilization in treating excess flows at aerated activated sludge plants is feasible and reliable.
2. Removal efficiencies of SS - 90%, BOD_5 - 80% and TOC - 79% can easily be maintained for excess flows.
3. Improved dry weather operation and performance can be realized at the existing plant.
4. The capital cost in dollars per unit of capacity can be lowered upon development of optimum variable operating ranges.

VI. Capability Profile: Storm Water Management Model

The Storm Water Management Model and other associated work completed and on-going in mathematical modeling are intended as prime tools for use in applying the total system concept in selecting alternatives for water pollution control. Work is continuing on providing additional capability and refinement of the SWMM and development of a family of models of varying degrees of sophistication for use by planners, engineers and decision makers.

This profile is extracted, with quotations, from a paper entitled "A Model for Assessing Impact of Stormwater Runoff and Combined Sewer Overflows and Evaluating Abatement Alternatives: by Harry C. Torno, U.S. EPA., Municipal

Pollution Control Division (RD-678), Washington, D.C. 20460. Copies of the model program may be obtained by sending a blank 9-track 1600 hpi computer tape to Mr. Torno and requesting the SWMM program. Program documentation will be furnished or is available separately.

A users group has been established to exchange information and experience on model applications and to act as a focal point for improvements. Further details are available from Mr. Torno.

The entire SWMM is large. There are over 10,000 FORTRAN statements and it requires a large computer for reasonable execution. For example, an IBM 360/65 with at least 512K bytes of core storage is required. Some users have executed the program on smaller computers (IBM 1130) by overlaying, reducing size and number of COMMON areas and by using portions of the program.

Using the rainfall (hyetograph) and system (catchment, conveyance, storage/treatment and receiving water) characteristics as inputs the SWMM determines quality and quantity of runoff, routes the runoff through a combined (or separate system) with specified storage and treatment facilities and operating policies, and thence into the receiving water, where impacts are identified. The output of the model consists of tables, hydrographs and pollutographs of BOD, suspended solids, dissolved oxygen and coliform. These can be displayed for selected points within the system as well as in the receiving waters.

The SWMM consists of 5 blocks, or groups of subroutines. They are:

A. EXECUTIVE Block - Provides control and service functions. All interfacing between the four computational blocks takes place through this block. The block includes a subroutine called COMBINE (a network aggretation routine) which allows collection of two or more output data sets, and allows combination of different data sets and manholes into a single data set with one manhole. This aids in modeling of large geographical areas.

B. RUNOFF Block - Computes the storm water runoff and its associated pollution loadings for a given storm for each subcatchment and stores the results in the form of hydrographs and pollutographs at the inlets to the main sewer system. Overland flow simulation is accomplished by a storage routing method using Manning's equation and the continuity equation. Overland flow does not begin until depression storages are full. Infiltration on previous areas is computed by Horton's exponential function. and is subtracted from water depth existing on the sub-catchment. Gutter flows are treated as a succession of steady-state flows, with outing accomplished using Manning's equation and the continuity equation. To use this block, the user must input the rainfall hyetograph and a discretization of the drainage basin into sub-basins of constant land from characteristics. The location and chara-teristics of the gutters and pipes also have to be described. In addition, the user must input street cleaning frequency and catchbasin data as well as the land use and other features of the different areas of the basin.

C. TRANSPORT Block - Routes flows through the sewer system. Prestorm conditions in the sewers are set up by computing dry-weather flow and infiltration and distributing them throughout the conveyance system. The Transport Block then routes the storm runoff (as determined by the RUNOFF Block), the dry weather flow (DWF), and the water that has infiltrated into the system through the main sewer pipes, and through a maximum of two optional "internal" storage tanks.

The routing scheme is based on a finite-difference solution of the St. Venant equations, in which normalized values of the flow and conduit cross-sectional area are used. When a pipe is flowing full and inflow exceeds outflow, the excess (surcharge) is stored at the upstream manhole. The flows are routed to a maximum of five outlet points. This block requires that the sewer system be discretized into pipe segments of constant size, slope, and type jointed by either manholes, control structures such as flow dividers, or "internal" storage tanks. An "internal" storage tank is described by its size, shape, outlet device, and unit cost. The outlet device can be either a pump specified to go on or off at a specified tank depth, a weir, or an orifice. The outlet device is used to specify the operation policy of the storage tank.

The DWF and quantity entering the sewer system are calculated by inputting to the model such parameters as daily and hourly pollution correction factors, land use and population of the subareas, and average market value of the dwellings in a subarea. If more exact data is available such as average BOD of flows, this can be used in place of some of the other data.

Infiltration is calculated by estimates of base dry weather infiltration and groundwater and rainwater infiltration, and such parameters as average joint distance. The use of the subroutines calculating DWF quality and quantity and infiltration is optional.

D. STORAGE Block - Simulates the changes in the hydrographs and pollutographs of the sewage as the sewage flows through one optional special wastewater treatment facility. The facility has to be located at one of the outfalls specified in the Transport Block. The treatment process is chosen by the user to consist of a sequence chosen from the following unit processes: "external" storage (same as "internal" storage except that it is located adjacent to an outlet of the sewer system), bar racks, fine screens, dissolved air flotation, sedimentation tanks, microstrainers, high rate filters, effluent screens, swirl concentrators and chlorinators and other chemical dispensers. The user can specify the sizes of the treatment processes or else can specify that the model is to select the sizes of the processes (except for "external" storage) such that a certain user-selected percentage of the peak flow receives treatment. The Storage Block also has the capability of calculating the capital, land, and operation and maintenance costs of the treatment processes chosen. The user has the option of either specifying the unit costs or using default values provided by the simulation model. The calculations in this block are based on the continuity equation.

E. RECEIVING WATER Block - Takes output from TRANSPORT or STORAGE and computes the impact of the discharges upon the quality of the receiving water. The receiving body of water is discretized by the user to consist of a network of nodes connected by channels. An option in the program allows two parallel channels to be used between junctions, to aid in simulating receiving bodies such as marshes. Each channel is of constant surface and cross-sectional area. Boundary conditions can be specified as a weir (outfall from a lake) or some tidal condition. The structure of the SWMM is schematically depicted in Figure 4.

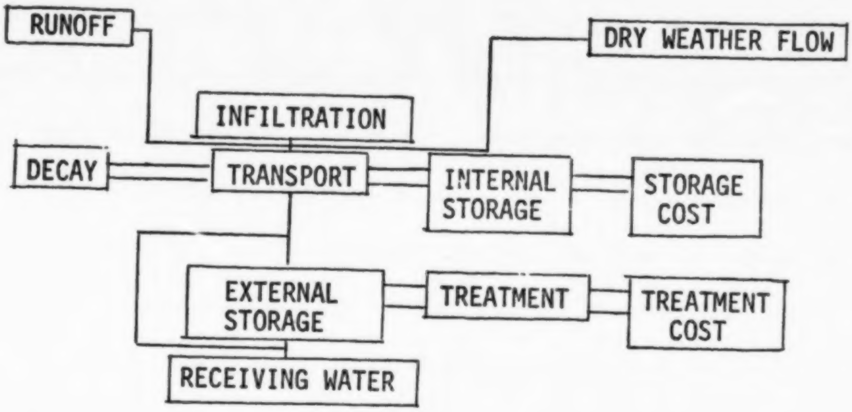


FIG. 4
SWMM Structure

The model was originally developed to consider only storm and combined sewer abatement alternatives. It has become apparent that the wet weather flows cannot be considered separately from the conveyance and treatment of sanitary sewage. This is brought into focus when considering the effects of various discharges (municipal treatment works, combined sewer overflows and industrial discharges) upon receiving waters. Modifications are now being made to include dry weather treatment facilities in the SWMM and to increase the capabilities for such pollutants as nitrogen, phosphorous, oil and grease.

Model verification and refinement has been accomplished in application to measured flow and analyzed water quality for catchments in San Francisco, Cincinnati, Washington, Philadelphia, and Lancaster, Pennsylvania.

Access to digital computers for execution of complex mathematical models is becoming more available to the engineer and such tools should be used in examining abatement alternatives to water pollution abatement.

VII. Summary

In the next three years there will be several EPA demonstration projects in which the evaluation phase will be completed. These projects were initiated as long as five years ago. Therefore, for the immediate future the program plan may best be described as an assimulation period. The data obtained from the full-scale facilities will be incorporated into data matrices relating to cost effectiveness for use in developing problem solution methods. Consequently, the start-up of full-scale demonstrations will be curtailed.

The following tables are based on the data collected for a report now in preparation* entitled "Urban Storm Water Management and Technology: An Assessment".

The tables are skeleton outlines and summaries of the cost and performance information collected which is pertinent to this paper.

Table 7

Estimated Costs of Sewer Separation

Regional Costs	Estimated cost[a], $/acre Type 2[b] (gravity)
New England	35,580
Middle Atlantic	24,350
South Atlantic	24,530
Southern	16,720
Midwest	10,710
West	9,250
National average	18,260

a. Adjusted to ENR = 2000.
b. Type 2 is constructing new storm sewers and using existing combined sewers for sanitary sewers.
Note: $/acre x 2.47 = $/hectare

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*Published as Report EPA-670/2-74-040, December 1974.

Table 8

Sewer Separation Versus Conceptual Alternatives

Location, (REF.)	Capital costs[a], $		Cost ratio[b]	Alternative
	Separation	Alternative		
Boston, Mass.	997,260,000	779,692,000	1:3	Deep tunnel storage
Bucyrus, Ohio	15,957,000	9,220,000	1:7	Lagoon system
Chicago, Ill.	6,772,255,000	1,322,378,000	5:1	Storage tunnels and quarries
Cleveland, Ohio	372,405,000	111,842,000	3:3	Offshore stabilization ponds
Detroit, Mich.	2,859,185,000	2,859,000	1,000:1[c]	Sewer monitoring & remote control of existing combined sewer storage system
Seattle, Wash.	15,486,000	8,185,000	1:9[d]	Computer controlled in-sewer storage system
Washington, D.C.	677,778,000	353,333,000	1:9	Tunnels & mined storage

[a] Adjusted to ENR = 2000.

[b] Ratio of separation cost to alternative cost.

[c] Alternative costs are for first phases only and do not include future total system.

[d] Separation costs are only for southwest and east central Seattle, while alternative costs are for the total combined sewer area.

Table 9

Summary of Storage Costs for Various Cities[a]

Location	Storage, mil gal.	Capital cost, $	Cost per acre, $/acre	Storage cost, $/gal.	Annual Operation & maintenance cost, $
Seattle, Wash.					
Control and monitoring system	--	3,500,000	--	--	--
Automated regulator station	--	3,900,000	--	--	--
	32.0	7,400,000	--	0.23	250,000
Minneapolis-St. Paul, Minn.	--	3,000,000	--	--	--
Chippewa Falls, Wis.					
Storage	2.8	744,000	8,260	0.26	--
Treatment	--	186,000	2,070	--	--
Jamaica Bay, New York City, N.Y.					
Basin	10.0	21,200,000	6,530	2.12	--
Basin and sewer	23.0	21,200,000	6,530	0.92	00
Humboldt Avenue, Milwaukee, Wis.	4.0	2,010,000	3,560	0.50	--
Boston, Mass.					
Cottage Farm Stormwater Treatment Station	1.3	6,200,000	--	4.74[b]	65,000
Chicago, Ill					
Storage and tunnels	5,570.0	1,323,000,000	5,500	0.24	--
Treatment	--	1,550,000,000	6,460	--	--
Storage basins	2,736.0	568,000,000	2,370	0.28	--
Collection, tunnel, and pumping[c]	2,834.0	755,000,000	3,150	0.27	--

a. ENR = 2000.
b. Includes pumping station, chlorination facilities, and outfall.
c. Includes 193.1 km (120 miles) of tunnels.
Note: $/acre x 2.47 = $/hectare, $/gal. x 0.264 = $/liter, mil gal. x 3.785 = Ml

Table 10

Cost of Sedimentation Facilities[a]

Location of facility	Size,[b] mgd	Capital cost, $/mgd	Annual operation and maintenance cost, $/mgd
Cambridge, Mass.			
Cottage Farm Stormwater Treatment Station	62.4	100,000[c]	1,240
Columbus, Ohio			
Whittier Street	192	32,000	--
Alum Creek	43	43,000	--
Milwaukee, Wis.			
Humboldt Avenue	192	10,500	--
New York, N.Y.			
Spring Creek-Jamaica Bay	480	44,000	--

a. ENR = 2000.
b. Maximum capacity assuming 30-minute detention time.
c. Includes pump station and screening facilities.

Note: mgd x 43.808 = l/sec

Table 11

Cost of Microstrainers and Ultrafine Screens for 25-Mgd Plants[a]

	Influent source	Loading rate, gpm/ sq ft	Capital cost[b]	Modified capital cost[c]	Operation and maintenance cost Annual	Operation and maintenance cost ¢/1,000 gal.
Microstrainers						
Philadelphia, Pa.	Storm overflow	25	$ 283,500	$283,500	$ 490	0.19
Taft Institute	Activated sludge effluent	16	382,800	245,000	1,480	0.57
Hypothetical	Sewage effluent	5-10	1,180,800	354,000	2,620	1.01
Chicago, Ill.	Activated sludge effluent	6.6	1,556,600	411,000	2,580	0.99
Ultrafine screens						
Fort Wayne, Ind.	Storm overflow	--	119,000[d]	--	--	--

a. ENR = 2000.
b. Installed.
c. Based on a loading rate of 25 gpm/sq ft.
d. Equipment cost only.
Note: ¢/1,000 gal. x 0.264 - ¢/1,000 liters
mgd x 43.808 = l/sec
gpm/sq ft x 0.679 = l/sec sq m

Table 12

Capital and Operation and Maintenance Costs for Biological Treatment[a]

Item	Contact stabilization, activated sludge	Trickling filter	Rotating biological contactor	Oxidation pond
Location	Kenosha, Wisconsin	New Providence, New Jersey	Milwaukee, Wisconsin	Shelbyville and Springfield, Illinois
Plant capacity, mgd	20	4	10.4	0.3 to 2.2[b]
Capital cost (construction cost, excluding land, $/mgd	78,300[c]	68,000[d]	30,000	43,800 to 55,000
Operation and Maintenance (annual cost assuming 250 hr/yr of operation), ¢/1,000 gal.	4.8	6.1[e]	4.4	1.0[f]

a. ENR = 2000.
b. Equivalent capacity which is a pond with a 10-day detention time.
c. Cost of pumps, aeration tanks, and final clarifier.
d. Includes cost of plastic media filter, final clarifier, side piping, and electrical work.
e. Approximate cost of dry weather flow.
f. Cost only for Springfield, Illinois.

Note: mgd x 43.8 = 1/sec
$/mgd x 0.0228 = $/1/sec
¢/1,000 gal. x 0.264 = ¢/1,000 1

Table 13

Estimated Capital and Operation and Maintenance
Costs for Typical Physical-Chemical Treatment Plant

Location	Capital costs, $			Operation and maintenance costs, ¢/1,000 gal.		
	10 mgd	25 mgd	100 mgd	10 mgd	25 mgd	100 mgd
Hypothetical CAST[a]						
Total	4,822,000	9,680,800	28,330,500	9.7	7.1	5.3
Annual	377,200	757,300	2,216,300			
Hypothetical PCT[b]						
Total	6,656,000	13,409,000	42,379,000	16.3	13.4	10.3
Annual	520,700	1,049,000	3,315,300			
South Lake Tahoe, Calif.						
Total	4,870,500	9,907,400	29,010,600	13.0	10.8	8.6
Annual	381,000	775,100	2,269,500			
Albany, N.Y.[c]						
Total	1,791,500	3,643,900	10,670,100	18.8	15.6	11.7
Annual	140,100	285,100	834,700			

a. CAST = conventional activated sludge treatment (for comparison only).
b. PCT = physical-chemical treatment.
c. Combined Sewer Overflow Treatment based on 100,000 gal/day pilot plant.
Note: mgd x 43.808 = 1/sec
¢/1,000 gal. x 0.264 = ¢/1,000 1

Table 14

Dissolved Air Flotation Cost for 25 Mgd[a]

Plant location	Fort Smith, Arkansas	Milwaukee, Wisconsin		Racine, Wisconsin	San Francisco, California	
	(Estimated cost)	Steel tank (Actual cost)	Concrete tank (Estimated cost)	(Estimated cost)	(Actual bid cost)	Average cost
Construction cost including pre-treatment devices[b]	$480,000	$580,000	$686,000	$830,000	$1,760,000	$867,000
Operation and maintenance						
Total cost, ¢/1,000 gal.	10.83	5.75	--	3.34	--	6.64
Chemical cost alone, ¢/1,000 gal.	--	4.17	--	2.71	--	5.11[c]

a. ENR = 2000.
b. Fort Smith used hydraulic cyclones for pretreatment; Milwaukee and Racine used 50-mesh ultrafine screens; San Francisco used only bar screens.
c. Seventy-seven percent of total operation and maintenance cost, which is the average percentage the chemicals cost at Milwaukee and Racine.

Note: ¢/1,000 gal. x 0.264 = ¢/1,000 liters

Table 16

Examples of System Operation Controls

Location	Operation scale	Operation controlled	Monitoring devices	Control devices
Seattle Wash.	Full	Overflow quantity	Level sensors, rain gages, wind gages, automatic samplers, telemetry units, computer, position sensors.	Gate regulators, tide gates, pumping stations.
Boston, Mass.	Full	Overflow quality	Level sensors, automatic sampler, Dall tube, residual chlorine analyzer.	Pumping station, sodium hypochlorite feeder, gate regulators.
Dallas, Tex.	Full	Sewer surcharge	Level sensor, magnetic flowmeter, temperature probe, computer.	Polymer injection feeder
San Francisco, Calif.	Full	Overflow quality	Magnetic flowmeters, level probes, differential pressure sensor.	Polyelectrolyte feeders, solids pump, bypass gates, dissolved air flotation units.
Mt. Clemens, Mich.	Full	Overflow quality	Magnetic flowmeters, level probes, automatic samplers, differential pressure sensors.	Pumping stations, aerated lagoons, microstrainer, pressure filters, chlorine-chlorine dioxide feeders.
Milwaukee, Wis.	Full	Overflow quantity	Level sensors, automatic samplers, magnetic flowmeters.	Mixers, chlorinators, bar screen, detention tank.
Minneapolis, Minn.	Full	Overflow quantity	Automatic samplers, rain gages, computer, pressure sensors.	Fabridam regulators, gate regulators.

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Table 15

High Rate, Deep Bed, Dual Media Filtration Cost Cleveland, Ohio, Pilot Facilities

Plant capacity[b]		Capital[a]	Operation and
cu m/min	mgd	cost, $	maintenance cost, $/yr
64	25	1,580,000	44,000
132	50	2,390,000	55,000
263	100	4,370,000	98,000
526	200	7,430,000	129,000

a. The cost data are based on an ENR of 2000.
b. Based on 5 hour overflow event with only temporary sludge storage. Sludge handling cost not included.